

BACKGROUND INFORMATION DOCUMENT FOR THE
DEVELOPMENT OF REGULATIONS TO CONTROL THE
BURNING OF HAZARDOUS WASTES IN BOILERS AND
INDUSTRIAL FURNACES - VOLUME II, INDUSTRIAL
FURNACES

Engineering-Science
Fairfax, VA

Jan 87



U.S. Department of Commerce
National Technical Information Service

NTIS

PB87-173837

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VOLUME II
INDUSTRIAL FURNACES

Submitted to

U.S. Environmental Protection Agency
Waste Treatment Branch (WH-565A)
401 M Street, S.W.
Washington, D.C. 20460

January 1987
66036.00/80A-D

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U.S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

Submitted by

Engineering-Science
Two Flint Hill
10521 Rosehaven Street
Fairfax, Virginia 22030

ENGINEERING-SCIENCE

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SECTION 1

SUMMARY AND CONCLUSIONS

One alternative disposal method for hazardous wastes in the United States is to burn the wastes in industrial furnaces. This report includes background information for use in developing and evaluating regulatory alternatives and in preparing a Regulatory Impact Analysis (RIA) of burning hazardous waste in industrial furnaces. Seven industrial furnaces were evaluated in depth; included were cement (wet and dry processes), lime, lightweight aggregate kilns, asphalt aggregate kilns, blast furnaces, spent acid recovery furnaces, and open hearth steel furnaces. The following characteristics make these furnaces likely candidates for disposal of hazardous wastes:

- o Process operations require high temperatures ($>1050^{\circ}\text{C}$)
- o Combustion gases exhibit long residence times (2-10 seconds)
- o Processes have the ability to burn low quality fuels
- o Efficient particle removal equipment already exists on most kilns
- o Alkaline raw materials used in production (or downstream operations in the case of spent acid recovery furnaces) tends to remove acidic gases from combustion products before discharge to the atmosphere

The owners and operators of furnaces used in the production of cement, lime, and lightweight aggregate, and asphaltic concrete do not produce liquid hazardous wastes. Instead, these wastes are generated elsewhere and transported to the furnace locations for destruction by combustion. Past waste burning in blast furnace was similar in that the wastes were generated off-site. The composition of hazardous wastes which might be burned in furnaces is highly variable depending on the generating source. Variable parameters include: Btu content, ash or solids content, water content, toxic metals content, chlorine content, and POHCs content.

In order to fire hazardous wastes efficiently in existing industrial furnaces, necessary modifications will generally include:

- o Installing waste storage and handling facilities
- o Installing equipment to characterize the waste
- o Installing equipment to pretreat the waste to improve its suitability for firing
- o Installing or modifying the burner gun for firing the waste
- o Modifying combustion controls
- o Installing waste feed rate, oxygen, and carbon monoxide monitors
- o Modifying air pollution control equipment

The number of furnaces burning hazardous wastes has not been established definitively. A recent survey found that 35 furnaces burn 86 million gallons of waste per year. The United States kiln population in these three industries totals about 800 kilns made up of about 275 cement kilns, 450 lime kilns, and about 75 lightweight aggregate kilns. These kilns probably offer the greatest potential for hazardous waste burning.

Results from the trial burns of eleven industrial furnaces were evaluated to determine their performance in burning hazardous wastes. These furnaces included: three dry process cement kilns, two wet process cement kilns, one lime kiln, one lightweight aggregate kiln, three asphalt aggregate kilns, one blast furnace. Testing procedures have not been standardized and quality assurance/quality control (QA/QC) procedures have not been established for testing of the destruction and removal efficiency (DRE) of hazardous wastes combusted in furnaces. A DRE of 99.99% for POHCs of hazardous wastes was achieved on all of the tests evaluated in this study. Of the furnaces evaluated by EPA, particle emissions were increased significantly in one cement kiln when the furnace was co-fired with hazardous waste. The reason may have been increased dust loading due to formation of salts or changes in dust resistivity which affected the efficiency of the electrostatic precipitator (ESP). Metal emissions tended to be somewhat proportional to the concentration in the waste feed. Hydrochloric acid (HCl) emissions from the tests were minimal even with high concentrations of chlorine in the waste feed. Removal is conjectured to take place in the furnace by reactions of chlorine with alkaline materials which end up in the product or in dusts collected by the air pollution control equipment. The lightweight aggregate kiln was controlled by a wet scrubber which removed HCl. Since lightweight aggregate does not react with chlorine like the clinker in cement or lime kilns, a scrubber would likely be required on an aggregate kiln burning hazardous waste to ensure that HCl is removed. PICs were not evident in the stack gas exhausts from the seven tested kilns, reflecting positively on the combustion efficiencies in the kilns.

Cost elements were provided for operating or retrofitting existing industrial furnaces. Capital and operating and maintenance costs were provided in 1982 dollars for those systems likely to be added at existing furnace operations if hazardous wastes were to be burned in the furnaces. This BID includes cost elements for:

- o Equipment to characterize the wastes
- o Equipment to pretreat the wastes by blending, straining, or thermal heating
- o New or modified burner guns to fire the waste
- o Equipment for combustion controls
- o Equipment for monitoring waste feed rates, oxygen, or carbon monoxide levels
- o Equipment for controlling particulate and gaseous stack emissions

EPA developed cost elements for waste storage and handling facilities as part of other regulatory development efforts.

SECTION 2

INTRODUCTION

The Resource Conservation and Recovery Act (RCRA) is one of the most important pieces of enabling legislation now being implemented by the U.S. Environmental Protection Agency (EPA). It requires EPA to develop regulations for the storage, handling, and disposal of hazardous wastes. It also requires that EPA encourage in its regulations the recycling of wastes. On May 19, 1980, EPA issued initial hazardous waste regulations which included requirements for hazardous waste incinerators. At that time, the burning of hazardous wastes in boilers and industrial furnaces was not regulated because EPA lacked sufficient information to determine the extent of the risk to public health posed by such burning as well as the extent of the regulatory controls that would be necessary to address the risks. Since 1980, EPA has researched the nature and extent of the practice of burning hazardous waste as fuel in boilers, and industrial furnaces. This document presents the results of research efforts for industrial furnace. A similar analysis for industrial boilers has been completed by EPA and is presented in Volume I of this document. The risk assessment analysis is presented in Volume III of this document.

OBJECTIVE

The objective of this project was to provide sufficient research and investigation into the concept and practice of burning hazardous wastes in industrial furnaces such that EPA could develop and evaluate regulatory alternatives. The information and data collected by this investigation constitutes a background information document (BID) for use by EPA in preparing its regulatory impact analysis (RIA).

SCOPE

The investigation was designed to characterize those industrial furnaces that have the potential to burn a hazardous waste as fuel. An industrial furnace is defined by 40 CFR 260.10 to mean any of the following enclosed devices that are integral components of manufacturing processes and that use controlled flame devices to accomplish recovery of materials or energy: cement kilns, lime kilns, aggregate kilns, phosphate kilns, coke ovens, blast furnaces, smelting, melting and refining furnaces (including pyrometallurgical devices such as cupolas, reverberatory furnaces, sintering machines, roasters, foundry furnaces), titanium dioxide chloride process

oxidation reactors, methane reforming furnaces, pulping liquor recovery furnaces, and spent acid recovery furnaces. These furnaces are typically high temperature devices (above 1500°F exhaust temperatures) used in industrial processes.

In addition to these specific furnaces, 40 CFR 260.10 set forth criteria for adding to the listing of industrial furnaces. The criteria are as follows:

- o The design and use of the device is primarily to accomplish recovery of material products.
- o The use of the device is to burn or reduce raw materials to make a material product.
- o The use of the device is to burn or reduce secondary materials as effective substitutes for raw materials, in processes using raw materials as principal feedstocks.
- o The use of the device is to burn or reduce secondary materials as ingredients in an industrial process to make a material product.
- o The use of the device in common industrial practice is to produce a material product.
- o Other factors as appropriate.

There are furnaces other than those listed in 40 CFR 260.10 that potentially meet these criteria. Glass melt furnaces, carbon black furnaces, and activated carbon retort kilns are examples of furnaces meeting the criterion of reducing raw materials to produce a product. Therefore, the universe of industrial furnaces is larger than those listed in 40 CFR 260.10. Additional furnaces can be added to the list of industrial furnaces under procedures identified in 40 CFR 260.20.

The quantity of hazardous waste currently being burned in industrial furnaces is not accurately known. In 1984, EPA conducted a mail questionnaire survey of industrial plants to estimate the quantity of waste being burned. Table 2.1 lists the quantity of waste burned in the industrial furnaces covered by the survey and estimates the total quantity of waste burned in all U.S. industrial furnaces using statistical scaling procedures. A total of 51 million gallons per year of hazardous waste was projected to be burned in industrial furnaces from the survey results. This estimate is known to be incorrect at this time since the burning of hazardous waste in blast furnaces has recently been discontinued. A survey conducted in April 1986 of the cement, lime, and lightweight aggregate industries found that 86 million gallons/year of hazardous waste was being burned in 35 kilns used in these industries (Reference 1).

Although limited, the available data indicate that appreciable quantities of hazardous waste is or has been burned in cement, lime, aggregate kilns, blast furnaces, and spent acid recovery furnaces. For this reason, these furnaces have been selected for an in-depth analysis to provide information needed to guide the evaluation of regulatory alternatives. Because

TABLE 2.1

LISTING OF INDUSTRIAL FURNACES FROM WESTAT SURVEY

Standard Indust. Code (#)	Industry	Device Type	Waste Type	Quantity Reported in Survey (gal/yr)	Projected National Quantity (gal/yr)
3312	Steel Products	Blast Furnace	- Tar sludge - Coke oven byproducts	208,000	395,000
3312	Steel Products	Blast Furnace	- Recovered solvents - Still bottoms - Other	5,861,729	9,379,000
	Same Facility	Blast Furnace	- Same	3,907,819	6,253,000
3312	Steel Products	Coke Ovens	- Tar sludge	8,163	15,500
3341	Secondary Smelting/ Refining Precious Metals	Film Recovery Incinerator	- Photo paper - Photo & x-ray film - Circuit boards - Photo fixing solutions	35,052 (316,500 lb)	3,502,000
2800	Chemicals and Allied Products	Fire Tube Boiler	- Allyl chloride wastes	914,514	3,577,000
		Recovery Unit	- Same - Heavy ends carbon tetrachloride - Distillation bottoms	1,470,115	
2819	Sulfuric Acid Manufacturing	Recovery Furnace	- Disulfide oils - Allyl alcohols - Acrylonitrile waste - Other	607,505	911,000

TABLE 2.1--Continued

Standard Indust. Code (#)	Industry	Device Type	Waste Type	Quantity Reported in Survey (gal/yr)	Projected National Quantity (gal/yr)
2819	Sulfuric Acid Manufacturing	Recovery/ Combustion Furnace	- Petroleum spent acid - Detergent sulfonation - Sulfonation acid - Etching acid	4,337,207 (36,200,000 lb/yr)	13,879,000
	Same Facility	Recovery Furnace	- Spent acids	(35,741,144 lb/yr)	
2600	Kraft Paper	Recovery Boiler	- Waste solvents	755	1,200
2631	Paperboard Mill	Recovery Lime Kiln	- Waste solvents	123,000	197,000
2631	Paperboard Mill	Recovery Lime Kiln	- Waste solvents	8,068,448	12,910,000
TOTAL				25,542,307	51,018,000

of their prevalence and importance as a industrial combustion device, reverberatory furnaces are also analyzed here. An open hearth furnace was selected to represent reverberatory furnaces.

The remainder of this document presents information on the selected furnace deemed necessary for EPA to conduct its RIA of alternatives for regulating the burning of hazardous waste in industrial furnaces. Although the furnaces subject to regulation that are not discussed here may differ somewhat from those selected, the information presented should be sufficiently applicable to them to provide an adequate technical basis for making regulatory decisions. Section 3 of this volume characterizes the selected furnaces in terms of their design, combustion methods, air pollution emissions and control techniques, and waste compatibility. Model furnaces are described in Section 3 to represent a "typical" unit of each type of furnace selected for the in-depth analysis. These models were developed for use in assessing the health risks associated with burning hazardous waste in industrial furnaces and for analyzing the economic impact of regulatory alternatives. The risk assessment results are discussed in Volume III to this BID and the economic impact analysis is documented in the RIA.

Section 4 describes the results of tests conducted to measure the performance of industrial furnaces in burning hazardous wastes. EPA funded tests of cement kilns, lime kilns, aggregate kilns, and blast furnaces to collect data for its development of regulations for industrial furnaces. Included in the discussion are: the DRE, the POHCs, emissions of particulate matter, HCl, PICs, metals and combustion gases.

Some of the cost factors needed for the economic impact analysis are presented in Section 5. Some costs such as those associated with storage facilities and permitting are not included because these are available from other sources. The cost items provided in this document fall into three categories: (1) conventional fuel prices, (2) capital costs to modify the furnace system to burn a hazardous waste, and (3) the major operating and maintenance cost associated with burning a hazardous waste.

REFERENCES

1. Stohrer, G. Burning Hazardous Waste in Cement Kilns. Report on a workshop held at Clean Site, Inc., on April 30, 1986 in Washington, D.C.

SECTION 3

CHARACTERIZATION OF INDUSTRIAL FURNACES

INTRODUCTION

This section presents the characterization of the industrial furnaces chosen in Section 2 for an in-depth analysis, i.e., cement kilns, lime kilns, lightweight aggregate kilns, blast, reverberatory, spent acid furnaces, and asphalt aggregate kilns. Each furnace will be described as to operation, combustion, compatible hazardous waste fuels, and existing air pollution control equipment. The intent is to describe important aspects to their firing hazardous waste fuels and to develop model furnaces for each of the selected industrial furnaces. These model furnaces will be used for the health risk assessment documented in Volume III of the BID.

PORTLAND CEMENT KILNS

General

The manufacture of Portland cement is one of the largest tonnage mineral commodities in the United States. Current (1982) United States annual production capacity (Reference 35) was 86,607,000 tons of clinker. This capacity was owned by 46 companies, with the top 10 having 59.4% of the installed capacity. These data and those which follow apply to clinker production (kiln capacity). Grinding capacity (finished cement production) data are very similar but reflect the existence of plants which only grind and blend to ship the finished product. There are a total of 135 plants producing clinker in 274 kilns, located throughout 38 of the 50 states. This dispersion reflects the industry's attempts to balance sources of raw materials and energy with the markets. The largest plant, National Gypsum's at Alpena, Michigan, has a clinker capacity of 2,450,000 tons per year (TPY). The largest kiln in the United States is Kaiser Cement's 5000 ton per day (TPD) unit at Permanente, California. The smallest plants, of which there are several, have clinker capacities of less than 150 TPD (<50,000 TPY).

Process Description

Portland cement consists of combinations of calcium, aluminum, iron oxides, and gypsum formed from a variety of raw materials. The basic chemical compounds of Portland cement are calcium silicates; primarily tricalcium silicate ($3\text{CaO} \cdot \text{SiO}_2$), referred to in the industry as C_3S . Dicalcium silicate ($2\text{CaO} \cdot \text{SiO}_2$), tricalcium aluminate ($3\text{CaO} \cdot \text{Al}_2\text{O}_3$), and tetracalciumalumina-ferite ($4\text{CaO} \cdot \text{Al}_2\text{O}_3 \cdot \text{Fe}_2\text{O}_3$) are the other major components in order of abundance.

Calcium sulfate (CaSO_4) and a variety of other compounds are added, usually in final grinding and blending, to provide special properties but the fundamental composition remains the same.

The calcium source is most often limestone, though other materials such as oyster shells and chalk are used. The cement rock limestone found in the Lehigh Valley of Pennsylvania is an example of a nearly perfectly balanced stone for cement manufacture but usually clay, shale, iron ore, and other materials must be added to provide the proper chemical formulation for the kiln reactions and the final product.

With a correct blend of raw materials, the mass is subjected to heat in a kiln or other unit where a number of physical and chemical processes occur. The temperature of the solid mass is raised from ambient temperature to about 1500°C (2700°F). The first occurrence is evaporation of free water at 100°C , followed by continued heating to about 500°C where combined water from clay is driven off. At about 900°C , calcination, or CO_2 evolution from carbonates, begins leaving the oxides. At this same temperature, other reactions begin such as that between lime and clay dehydration products. The material reaches the point of incipient fusion at about 1250°C and cementitious product (clinker) begins to form with the clinker reaching a maximum temperature of about 1500°C , which is required to form the tricalcium silicate, a requirement for quality cement.

Although some of the intermediate chemical reactions are exothermic, the overall chemical process requires the addition of about 1.8×10^6 Btu of heat per ton of product (Reference 1). This heat is supplied at the rate of 3 to 6×10^6 Btu/ton by the combustion gases from burning fossil fuel, primarily coal. Fuel use in cement kilns has followed the national trend of conversion from gas and/or oil to coal. Currently, 96% of the total installed capacity uses coal as primary fuel compared to 30% in 1974 (Reference 2).

Figure 3.1 (from Reference 1) portrays the basic components of a Portland cement plant. The wet and dry processes differ in the use of wet and dry grinding and in the kiln configuration. The grinding operations are not of much concern for this analysis, but the kiln configurations are of great interest. Wet process plants use "long" kilns (450-600 feet) to obtain the necessary heat transfer for water evaporation. Dry process plants can have a long kiln, medium kiln, or short kiln (200 feet), depending on whether a preheater and/or precalciner is used. The problems with burning hazardous waste in precalciner/preheater kilns are described in the dry process portion of this section.

The kilns are operated with solids flow countercurrent to the combustion gases, with the solids traveling down the slightly ($3-8^\circ$) inclined rotating kiln. The exhaust gases exit the kiln at temperatures of $450-700^\circ\text{C}$ depending on kiln type and energy conservation practices. The gases typically contain from 6% to 30% of the solids feed rate as dust, 20% by volume water vapor (wet kiln only) and up to 30% CO_2 , along with 10-1000 ppm CO, 10-2000 ppm SO_2 , and 100-1500 ppm NO_x . The gases are transported to the pollution control equipment with an induced draft (ID) fan.

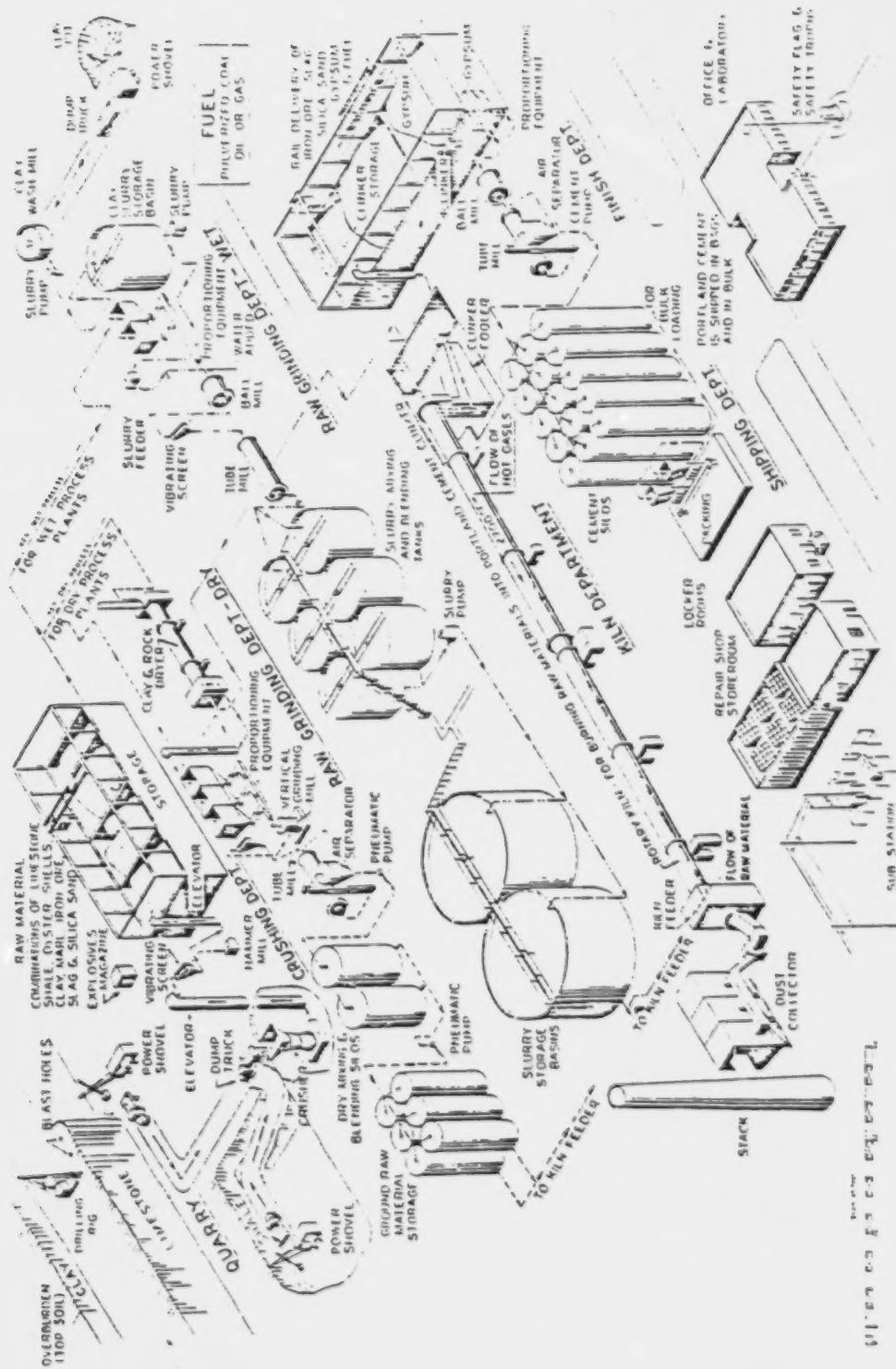


FIGURE 3.1 - Isometric Flowchart for the Manufacture of Portland Cement by both Dry and Wet Processes (Reference 1).

Air pollution control equipment in use at cement plants encompasses most typical control systems. Wet scrubbers, not surprisingly, are not common and are found only on raw material handling systems, as in quarries. Finish grinding, storage, and cement transfer operations are controlled almost exclusively with fabric filters. Kiln emissions are controlled by fabric filters or ESPs, often preceded by multiclones. Clinker cooler emissions are treated by fabric filters, ESPs, gravel bed scrubbers, and cyclone separators. The products of hazardous waste combustion (complete or incomplete) will appear in the clinker, in the kiln dust, and in the kiln exhaust gas. That which appears in the dust and kiln gas as a potential discharge is of much greater concern than that which appears in the product. One of the advantages of kiln disposal is the potential for trace metal removal as part of the product.

Portland cement was first made by the wet process. Recent developments in dry grinding and blending technology and the increased cost of energy have made the dry process more economical. New construction since 1971 has been almost exclusively dry process kilns (Reference 3). At the end of 1982, though, 38.8% of the installed capacity used wet process kilns (Reference 3).

The raw materials are the same for both processes, but the sequences and operations for raw material blending, grinding, and crushing are different. In the wet process, the raw material is crushed and proportioned to the approximate desired raw mix constitution, and then slurried with water to about 60-70% solids. The slurry is ground in wet mills (ball, tube/ball combination) to final size specifications and pumped to storage tanks. Fine adjustments of the mix may be made in the tanks before pumping to the kiln.

The kiln length and heat requirements are the major differences between the wet and dry processes. The first 20-25% (100-150 feet) of a wet process kiln is devoted to evaporating the water used for slurring. This results in added water content and lower temperature of the exhaust gas compared to dry process kilns. Generally, wet process kilns require 5.5 to 6.0×10^6 Btu per ton of clinker produced, while dry kilns require about 4.5 to 5.0. Most wet kilns are equipped with a "chain" section in the last (nearest the feed end) 60-100 feet of the kiln as an energy conservation measure. The chains serve to increase the heat transfer rate from the discharge gas to the raw feed and to aid in keeping the mass moving down the kiln. Solids residence time in a wet kiln is usually on the order of two hours, with gas residence time of about 6-10 seconds. However, the important factor is not total residence time but, the time/temperature profile. In this regard, both types of kilns have the same basic profile for the combustion and clinkering zones.

Referring to the solid/gas temperature profiles of Figure 3.2, the fuel combustion zone is located in the first 50 feet of the kiln. Burner arrangements place the centerline of the flame on the cylinder axis of the kiln. The gases have cooled to less than 2000°F at about 40% of the kiln length for a wet kiln. For an average kiln length of 525 feet (Reference 1) the effective incineration zone is about 200 feet. In general, this is consistent with the dry process (preheater/precalciner) kiln design where final calcination and clinkering occurs within this same temperature range (2000 to 2700°F), and requires a similar length of rotary kiln.

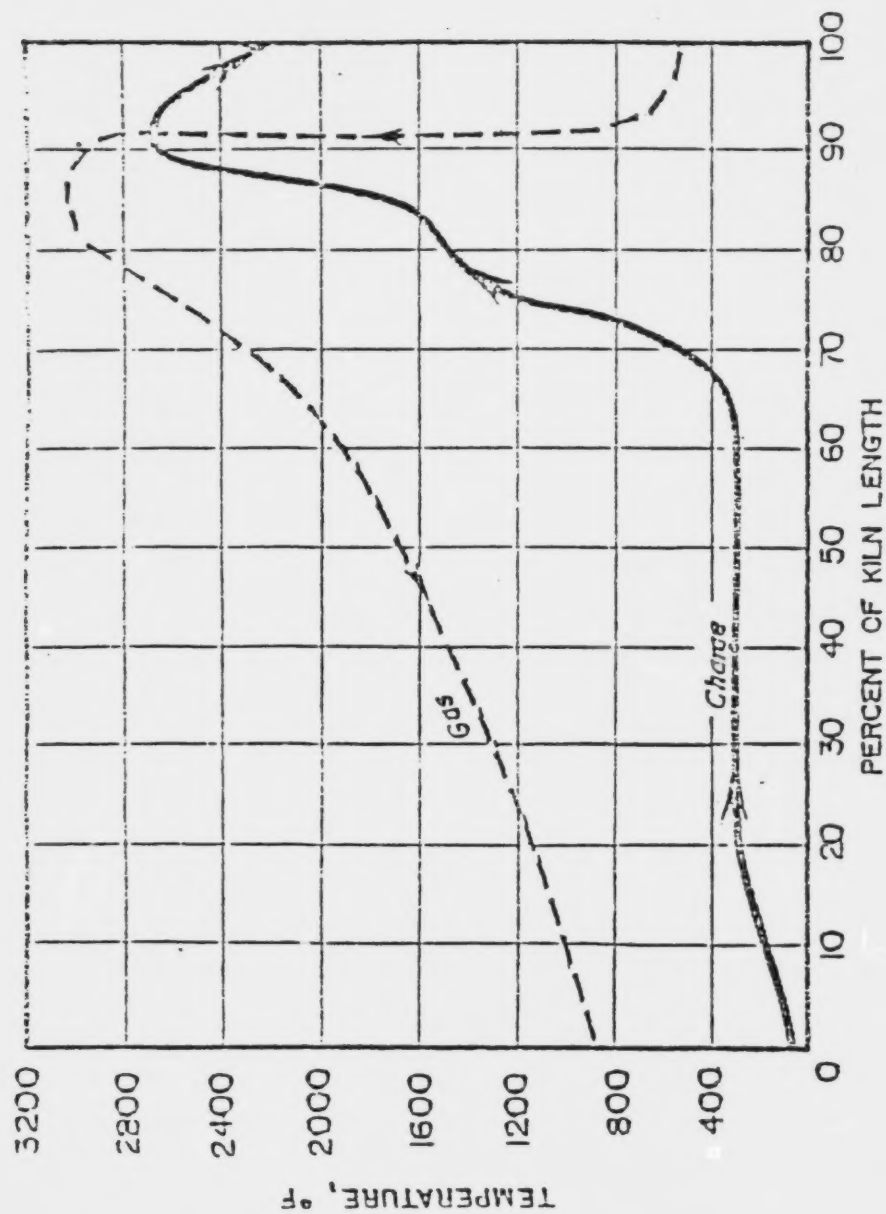


FIGURE 3.2 - Typical Wet Kiln Temperature Profile (Reference 3).

NOTE: Combustion Gases Flow Countercurrent to the Cement Charge

The average wet process kiln has a capacity to produce 266,500 tons of clinker per year, or about 850 TPD. The kilns normally operate 24 hours per day and seven days per week, but average 310 days per year. The largest can produce over 4000 TPD while the smallest produces no more than 220 TPD. The discussion later in this section describes a representative wet kiln, recognizing that this representative kiln probably does not really exist.

The hardware used to dry, calcine, and clinker the raw material is the difference between the wet and dry processes. The only inherent process difference is that less moisture needs to be evaporated in a dry process kiln allowing kilns to be shorter than equivalent capacity wet kilns. Kiln configurations take three basic forms:

- o Rotary kiln only
- o Rotary kiln with preheater
- o Rotary kiln with preheater and precalciner

While the physical and chemical processes are the same for all three configurations, the addition of a preheater with or without precalciner improves the process efficiency, through fuel savings and economies of larger production units.

A preheater system, using kiln exhaust gas, is applied to raise the solids temperature and partially calcine the charge before feeding it to the rotary kiln. The preheater accomplishes the same tasks as the "back end" of a long dry kiln but more efficiently. The kiln is one-half to two-thirds shorter if a preheater is used.

Two types of preheaters are in use today; the traveling grate type and, more commonly, the suspension type. Traveling grate preheaters operate in much the same manner as clinker coolers or traveling grate stoker boilers. Kiln exhaust gases are pulled through a bed of pelletized raw mix supported on a traveling grate. The suspension type of preheater consists of from one to as many as four cyclones arranged in a vertical tower with solids flowing down countercurrent to kiln exhaust gases. A graphic representation of a 4-stage preheater kiln is shown in Figure 3.3a. Figure 3.3b shows a 4-stage unit with a precalciner.

The third major configuration adds a precalciner between the back end of the kiln and the preheater tower. The precalciner is fueled separately from the kiln and may use up to 60% of the total fuel burned. Up to 95% of the charge may be calcined in the preheater/precalciner and this means of precalcination results in a uniform mass being fed to the kiln for final calcination and clinkering. Thus, the kiln operations can be optimized for the clinkering reactions, and the preheater/precalciner operated to optimize CO₂ evolution. However, the direct contact between hot kiln gases and raw feed in the preheater and the precalciner can result in condensation of sulfur and alkalies on the feed and increase the concentration of these substances in the clinker enough to make products of unacceptable quality. This direct contact and resultant condensation would become even more problematic if waste fuel were burned in the precalciner. Problems discussed below for kiln combustion of waste fuel in preheater kilns would be magnified. Further, the time/temperature profile of the precalciner flame is much less conducive to complete destruction of organics. Residence time is shorter and direct gas-to-solid

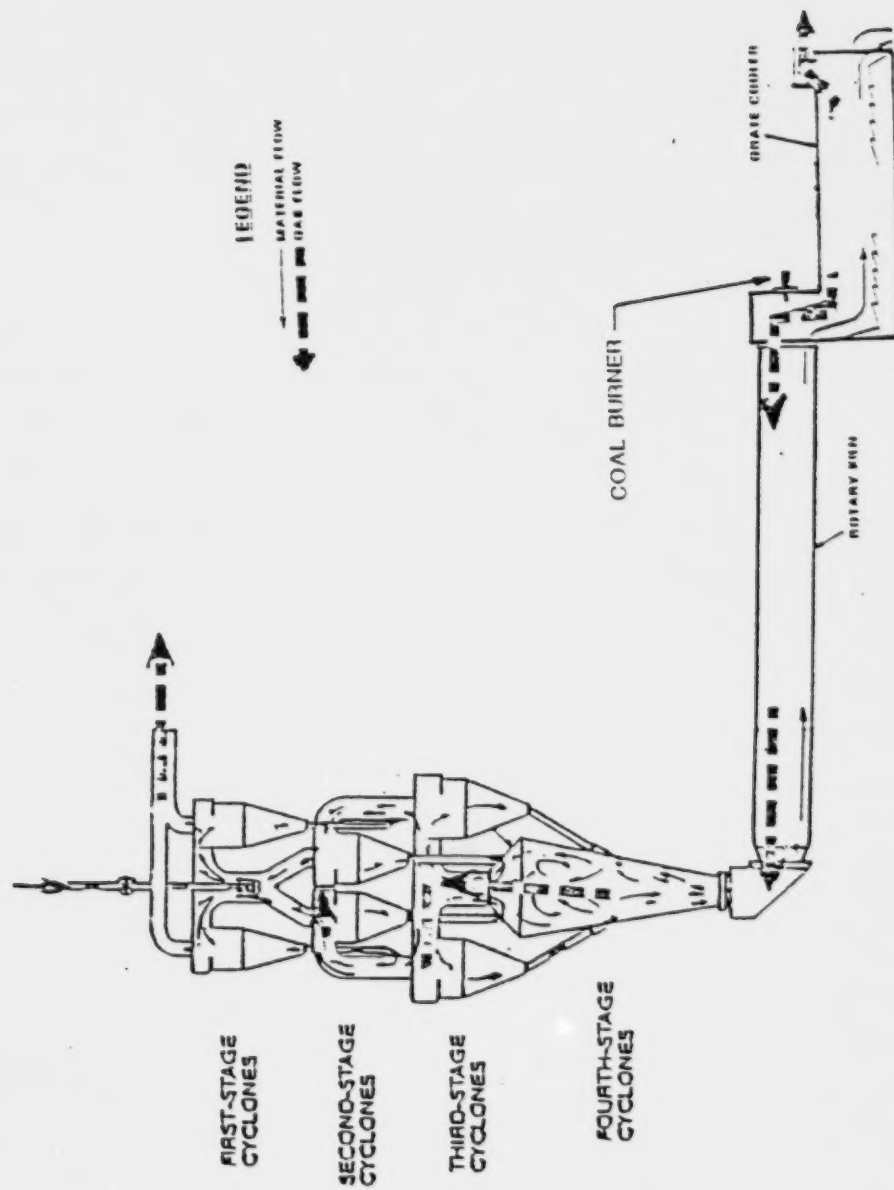


Figure 3.3a - SCHEMATIC DIAGRAM OF 4-STAGE SUSPENSION PREHEATER

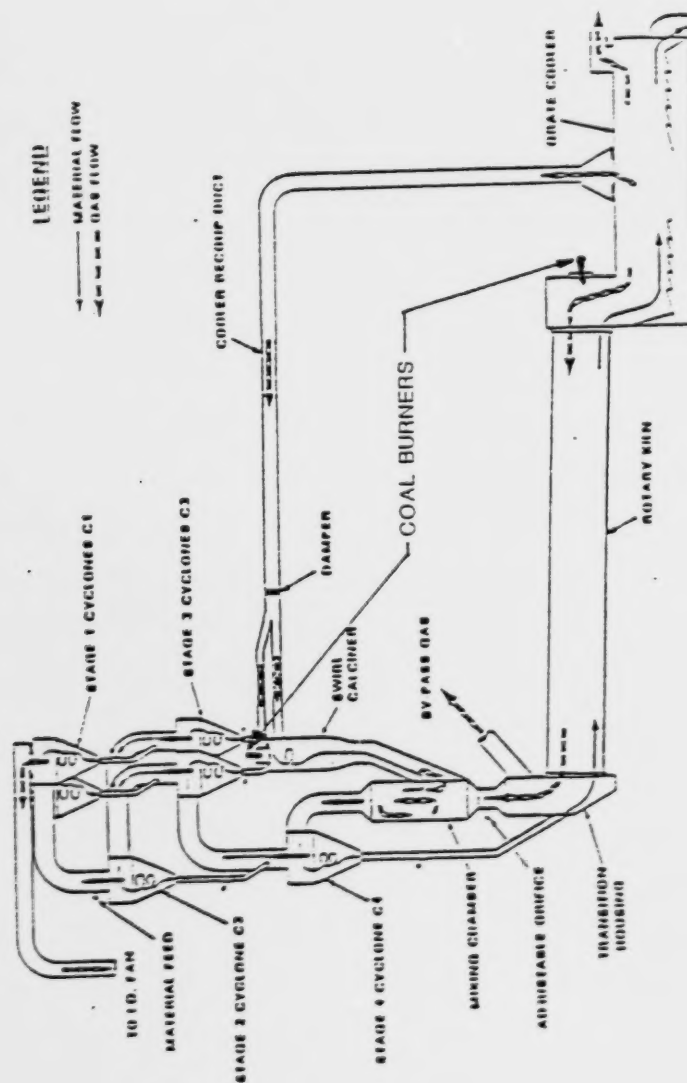


Figure 3.3b - SCHEMATIC DIAGRAM OF 4-STAGE SUSPENSION PREHEATER WITH PRECALCINER

charge heat transfer cools the gas stream much more rapidly than the gas-to-kiln-to-solid charge transfer in a rotary kiln.

Alkali concentration in the clinker also occurs in wet and long dry kilns but to a lesser degree because more heat transfer occurs from kiln walls to solids than gas to solids. The alkalies may also condense in the gas stream, but they do not deposit on the feed particles. The condensation problem becomes more acute if chlorinated fuels are burned since the alkali metal chlorides condense even more readily than the oxides and can create kiln lining deterioration, preheater plugging, and increase the alkali recirculation in the system. One solution to the problem of alkali buildup commonly practiced is bypassing some of the kiln exhaust to a separate control device, thus removing some of the more contaminated feed.

Overall, dry process kilns have an average capacity of 358,000 tons of clinker/year, or 1090 TPD. Long rotary kilns only average 236,000 TPY, while preheater kilns average 394,000 TPY, and preheater/precalciner kilns average 855,600 TPY (Reference 3). The capacity data reflect kiln age and advances in technology as the newer kilns are larger and employ the most sophisticated process technology. For example, the 20 precalciner kilns of the 274 operating kilns accounted for 18.8% of the total installed capacity in 1982, whereas there were no precalciner kilns in 1971.

The other unit operations of the cement manufacturing process have little relevance to the destruction of hazardous waste. The clinker exits the kiln to coolers where the temperature is reduced by pushing or pulling ambient air through it. Some preliminary size reduction is done before conveying the clinker to interim storage, and then to final grinding and formula adjustment.

Combustion

Nearly all kilns use coal as primary fuel. In general, lump or crushed coal is delivered to the plant and stored in piles. It is transferred from these piles to a silo or storage bin holding a 24-hour supply. From the silo or storage bin, the coal is fed to a pulverizer and carried from the pulverizer to the burner with ambient air preheated by mixing with exhaust air taken from the clinker cooler. A system of dampers is provided to achieve the desired mixing of the ambient air and hot air from the cooler. Secondary-combustion air may be added at the burner pipe, which is a large (approximately 24-inch diameter) tube oriented on the rotational axis of the kiln. Auxiliary fuel burners for startup and emergency are located along or inside the coal pipe. Gas and oil are the usual supplemental fuels, but wastes may be used. Common practice, usually because of permit conditions, is to use conventional fossil fuels rather than hazardous wastes for such occasions. The ash contribution from coal to the total loading of dust is considerably less than that from the kiln solids. Most operators practice readdition of collected dust to the kiln in one form or another. A typical coal with 8% ash burned in a wet kiln at 5.5×10^6 Btu/ton of clinker produced will generate 40 pounds of ash per ton of clinker produced, compared to 100-200 pounds of kiln dust per ton, plus insufflation (pneumatic reinjection of collected dust) of 100 pounds per ton.

Emission and Air Pollution Control

The exhaust gas leaves the kiln at temperatures ranging from 400 to 525°F, carrying the coal ash, oxides of sulfur from fuel and raw material, nitrogen oxides from the combustion process, and kiln dust. The kiln dust consists of particles of raw mix, partly or completely calcined matter, and clinker particles in that approximate order of abundance. The gas also contains the water driven off in the kiln as well as any unburned fuel, and any volatile organic contaminants present in the kiln solids or slurry water.

The control equipment in use on a wet-process kiln is nearly always a fabric filter or an ESP, often preceded by a centrifugal separator (cyclone or multiclone). Table 3.1 summarizes the control equipment used on wet process kilns and their collection efficiencies as reported in the EPA NEDS data base. This 1980 data base may include plants that are no longer operating. However, the basic population should be approximately the same except that baghouses probably now comprise a larger percentage, because more ESPs tend to have been installed on the older plants.

Table 3.2 summarizes some of the pertinent operating characteristics of the various dry kiln configurations. The exhaust gas temperatures reflect the gas leaving the system. Except for the 4-stage unit, cooling is required before the gases are ducted to control equipment. Normally, the temperature limit for ESPs is 700°F while it is 500°F for baghouses.

Control equipment in use on dry process kilns encompasses the same types as used for wet kilns. Table 3.3 shows the use of these systems on dry kilns. The data source does not permit identification of how these facilities are divided among the different kiln configurations; however, application of fabric filters is a fairly recent development in the history of cement manufacture and one would expect that a high percentage of these devices are installed on the newer kilns with preheater/or preheater and precalciner together.

Compatible Wastes

The same characteristics that permitted the industry to easily convert to coal fuel are potential advantages to use of hazardous waste as fuel or at least to recover the heating value of hazardous wastes. The nature of the process necessitates that the heat source (combustion gas) be hotter than the solid mass. Thus, the combustion gas must be hotter than the 1500°C needed to form cement. The conclusion to be drawn from these chemical and thermodynamic interactions is that in order to manufacture high quality cement, the maximum combustion gas temperature must be around 1700-1800°C. At this temperature, complete combustion (given sufficient oxygen and time) is virtually certain, especially for the typical components of a waste fuel. Considering that the gas residence time in a long wet kiln is on the order of 10 seconds, the temperature profile of Figure 3.2 shows that the combustion gases are above 1000°C (2000°F) for several seconds during its passage through the first 40% of the kiln length. The temperature profile in a dry process kiln is similar, but without the initial heat-up period during the last 20 to 60% of kiln length.

TABLE 3.1

AIR POLLUTION CONTROL DEVICES INSTALLED ON WET PROCESS CEMENT KILNS^a

System Type	Number of Kilns	Percent of Total	Average Reported Efficiency
Electrostatic Precipitator	73	51.0	98.2%
Electrostatic Precipitator + Centrifugal Separator	14	9.8	98.4%
Fabric Filter	27	18.9	99.4%
Fabric Filter + Centrifugal Separator	18	12.6	99.5%
Other	<u>11</u>	<u>7.7</u>	<u>92.4%</u>
Subtotal	143	100.0	N/A
No Data Reported	25		

^a Taken from the 1980 EPA National Emissions Data System (NEDS).

TABLE 3.2

OPERATING CHARACTERISTICS OF DRY PROCESS CEMENT KILNS

Kiln Type	Back End Temp. (°F)	Fuel Requirement (MMBtu/ton)	Dust Loss (% of feed)
Wet	450-550	5.5	10-15
Long Dry	1300-1400	4.2	25-30
Insulated Long Dry	800-900	3.4	30-35
1-stage Preheater ^a	750-800	3.3	10-12
2-stage Preheater ^a	750-800	3.2	8-10
4-stage Preheater ^a	650-680	2.8	6-8
4-stage Preheater ^a with Precalciner	600-650	2.8	5-6

^a Temperatures are preheater exhaust; rotary kiln exhaust gas is 1700-2000°F for all types of preheaters.

TABLE 3.3

AIR POLLUTION CONTROL DEVICES INSTALLED ON DRY PROCESS CEMENT KILNS^a

System Type	Number of Kilns	Percent of Total	Average Reported Efficiency
Electrostatic Precipitator	39	29.8	98.6%
Electrostatic Precipitator + Centrifugal Separator	11	8.4	98.1%
Fabric Filter	42	32.1	99.4%
Fabric Filter + Centrifugal Separator	26	19.9	99.8%
Other	<u>13</u>	<u>9.9</u>	<u>94.0%</u>
Subtotal	131	100.0	N/A
No Data Reported	36		

^a Taken from the 1980 EPA National Emissions Data System (NEDS).

Several operators in the United States and abroad have burned hazardous waste in their kilns, both as supplemental fuel and as a means of waste disposal, and have collected data on the effect of these practices. In a previous review for EPA, A.T. Kearney, Inc., (Reference 4), discussed the potential effects of several likely components of waste fuel on cement production. There are several elements which affect the quality of the product (e.g., sodium, potassium, magnesium) but which are not usually present in hazardous wastes in sufficient quantities relative to the feed to be significant. There are other elements found in wastes in great enough proportions to radically affect kiln operations and prime among these is chlorine. There are a number of trace metals commonly found in hazardous wastes such as lead, vanadium, beryllium, mercury, chromium, cadmium, which do not markedly affect the kiln operations or cement quality, but which could have an impact on the environment.

Test data collected to date support the hypothesis that the process conditions required for the production of quality cement will result in the complete destruction of hazardous waste. However, even the best kiln operation will produce poor quality cement at times, so the problem of ensuring adequate destruction of wastes becomes one of defining an appropriate set of conditions of kiln operations. Two aspects of waste destruction need to be considered:

- o Destruction combustion or removal of organic constituents
- o Collection and removal of inorganic constituents

The DRE of a kiln involves fuel combustion variables for organics while trace contaminant removal involves kiln chemistry and pollution control equipment operations. In practice, both are interrelated as combustion affects kiln chemistry, and to a lesser extent, kiln chemistry affects combustion.

Determination of those kiln operations which could relate to achievement of some regulatory standard (e.g., 99.99% DRE or 99% HCl removal) focused on the types of process data routinely collected by kiln operators. The parameters appearing to have the most direct connection to combustion efficiency and therefore DRE, are kiln gas oxygen and carbon monoxide concentrations both of which are usually monitored with continuous analyzers at existing kilns. The oxygen measurements are used to ensure an oxidizing atmosphere both for clinker quality and combustion purposes. The carbon monoxide monitors are typically used as "combustibles" monitors to prevent explosions or fires in emission control systems, or to provide evidence of overall reducing conditions in the kiln, unburned fuel, and unburned hydrocarbons entering the kiln in the raw materials.

A direct measurement of maximum clinker temperature might provide assurances of sufficiently hot combustion gases as well as proper kiln operations. However, at present there is no suitable technique for direct measurement of clinker temperature.

Waste fuels have been burned in some types of dry process kilns with varying degrees of success from a process viewpoint. It has been reported that many cement manufacturers have conducted short-term experiments with waste fuels, but none of their results have been reported. Test results have been reported for one trial of chlorinated waste and this was at St. Lawrence Cement's Mississauga plant in Canada. When burning chlorinated

waste in their kiln with a 4-stage preheater (no calciner), the alkali bypass duct continually plugged and stack particle emissions increased dramatically because of the control equipment's inability to handle the increased dust load. The tests were conducted using a fuel containing 40% chlorine by weight (trichloroethylene is 89% chlorine by weight) and resulted in a total chloride input to the kiln amounting to 0.8% of the clinker rate. The fuel replacement rate was 12%, but only 65% of the waste fuels heating value was realized in the kiln as the rest went to vaporizing metal chlorides (Reference 4). Other tests have been performed on long dry kilns in this and other countries with better success. No data are available on tests in preheater/precalciner kilns. Currently, five dry plants burn waste fuel and all are long kilns.

LIME PRODUCTION

General

Lime is one of the oldest manufactured chemicals known and the term applies to a variety of calcium compounds in the oxide and hydroxide forms. Over 20,000,000 tons of lime were produced in 1979; 36% of this was used for steel making (basic oxygen furnace), 8% for water purification, 6.5% for alkalies, and 5.5% for pulp and paper manufacture (Reference 5). The product derives from limestone containing varying amounts of magnesium in addition to the calcium; high magnesia limes are referred to as dolomitic. The oxides are called quick limes and the hydroxides are known as slaked (slack) limes.

The calcination of limestone to lime is performed in rotary kilns, fluidized bed kilns, vertical shaft kilns, and rotary hearth kilns, but 90% of the production comes from the horizontal rotary kilns (Reference 6). There are nearly 60 companies producing lime at 155 plants in this country, not including such captive facilities as lime regeneration in the pulp and paper industry. These 155 plants, located in 41 states and Puerto Rico, contain over 450 kilns, with the top 10 manufacturers producing just under 50% of the total output (Reference 4).

The rotary kiln process is the only one discussed in this document. As previously identified, horizontal rotary kilns account for 90% of the production; each of the other types are suitable only for a limited scope of raw materials; operating conditions of lime rotary kilns are similar to cement kilns; and operations of lime kilns appear compatible with hazardous waste incineration.

Process Description

Calcination of limestone in horizontal rotary kilns occurs in the same manner as described for Portland cement production because the raw materials are essentially the same (limestone) and the process is equivalent (dissociation of CaCO_3 to CaO and CO_2) except that CaO is an intermediate in cement production. Most lime kilns employ crushed or ground limestone as feed so water evaporation requirements are minimal (comparable to dry process cement). The charge is heated to evaporate any free moisture, which is usually less than 1% and then to dehydrate the feed. High-calcium limestone begins to dissociate at 1650°F while dolomitic limestones dissociate over a range of 1340 to 1650°F. The kilns are operated at maximum temperatures of 1800 to

2300°F to provide acceptable heat transfer rates, and at high gas flow rates to rapidly remove the evolved CO₂ to prevent recarbonation. Fuel requirements are anywhere from 4 to 10 x 10⁶ Btu/ton of lime (Reference 6) to meet the required heat of reaction of 2.8 x 10⁶ Btu/ton. On the average, the rotary kilns use about 7 x 10⁶ Btu/ton (Reference 7) plus the energy needed for kiln rotation, material handling, and pollution control equipment.

Kiln production rates range from 250 to 2,500 TPD, with the average plant producing approximately 135,000 TPY (about 450 TPD). The largest is Marblehead's Thorton plant which is rated at 600,000 TPY or 1,950 TPD (Reference 4). Kiln sizes are comparable to cement kilns though somewhat shorter and range from 6 to 14 feet in diameter and from 60 to 450 feet long (Reference 5). Fuel is burned in typical rotary kiln manner with a burner or burner pipe pointed up the kiln's rotational axis. Kiln slope is 3-5° which is a bit lower than that for cement kiln (6°).

Combustion

Coal accounts for almost 70% of the fuel used in lime production while natural gas is used for 23%. Oil, liquified petroleum gas (LPG), electricity, and other fuels comprise the remainder.

Emissions and Air Pollution Control

Kiln gases are exhausted at temperatures ranging from 500 to 1400°F. Emissions are controlled with fabric filters, ESPs, venturi scrubbers, and occasionally, gravel bed filters. Gas cooling required for baghouses and ESPs is provided by stone preheaters, water evaporation, ambient air dilution, or radiation/convection heat exchangers.

Control of emissions of particulate matter has been the major concern with respect to lime kilns, though they also emit sulfur oxides, nitrogen oxides, and carbon monoxide. Typically, kiln exhaust gases contain 50% of the fuel sulfur if high sulfur fuel is burned, resulting in an SO₂ concentration of 1500-2000 ppm SO₂. The concentration of NO_x is about 50-200 ppm (average 160 ppm) and of CO is 10-150 ppm (average 110 ppm). Wet scrubbers result in an alkaline scrubbing medium if recirculation is practiced and will remove HCl to some degree if chlorinated fuel is burned. Some acid gas adsorption by the reaction of the gases with lime particles suspended in the gas stream is also likely.

LIGHTWEIGHT AGGREGATE KILNS

General

The term "lightweight aggregate" describes a range of special use aggregates that have an apparent specific gravity considerably below normal sand and gravel. These lightweight aggregates will range from extremely light materials used for insulative and non-structural concrete all the way to expanded clay-like materials used for structural concrete. Some of these, e.g., pumice and scoria, occur as natural aggregates while the others are produced

by a variety of processes. Of interest to this study are the expanded clay-like materials (i.e., clay, shale, or slate), produced by the rotary kiln sintering process. The kilns used in producing them are very similar to those used in the Portland cement and lime industries. This class of lightweight aggregate is used in concrete in place of sand and gravel or stone. Concrete made with this aggregate has about the same strength but weighs approximately one-third less than concrete made with sand and rock.

There are 30 plants in operation in 24 states producing lightweight aggregate by the rotary kiln process (Reference 7). These plants typically have two or three kilns (Reference 8). Thus, there should be between 60 to 90 of these kilns in the United States. The 1979 total capacity of the rotary kiln process was approximately 15×10^6 cubic yards of product (12.6×10^6 tons) (Reference 2). Six plants have ceased operation since 1979 and therefore the current capacity is somewhat lower.

Process Description

A typical plant consists of: (1) quarries from which the raw material (clay, shale, or slate) is extracted; (2) raw material storage and handling facilities; (3) one or more rotary kilns to convert the clay-like material into lightweight aggregate; (4) aggregate product handling and storage areas; (5) fuel storage and handling facilities, (6) a kiln burner system for coal and/or liquid fuel; and (7) a kiln emissions control system with a wet scrubber being the most common type of control equipment.

The raw material used is a highly siliceous clay, shale, or slate that exhibits a bloating characteristic when heated. The bloating results because certain minerals in the raw material liberate gases at the temperature of incipient fusion. The raw material is bloated to the point that it becomes soft and pliable, but not to the point of completely melting.

In the rotary kiln process (Figure 3.4), the raw material is crushed and introduced at the upper end of a kiln similar to the type used in the lime industry. (In fact, some aggregate kilns were formerly lime kilns.) In passing through the kiln, the material reaches a temperature of 1900° to 2100°F, and begins to become plastic. Internal gases causes the material to expand, or bloat, and create a mass of small, unconnected air cells, which are retained after the material cools and solidifies. After leaving the kiln, the material is cooled and then crushed and graded.

As with cement and lime production, maintaining the proper temperature in the kiln is crucial to making good lightweight aggregate. Heat for the process is provided by a burner located at the clinker discharge (lower) end of the kiln. The temperature of the kiln is controlled by adjusting the fuel feed rate, kiln rotation speed, and amount of air flow through the kiln. Control is almost exclusively manual. The average temperature in the hottest zone of the kiln (near the burner) is usually held in the range of about 1900° to 2100°F (i.e., the temperature of incipient fusion of the charge material), which is nearly identical to the temperature range at the hottest zone of a lime kiln. If the kiln temperature falls below that general range, the clay structure is not converted to the desired inorganic formation. If the temperature rises above the desired range, the material fuses, forming a heavy coating on the kiln interior that must be removed manually after stopping and

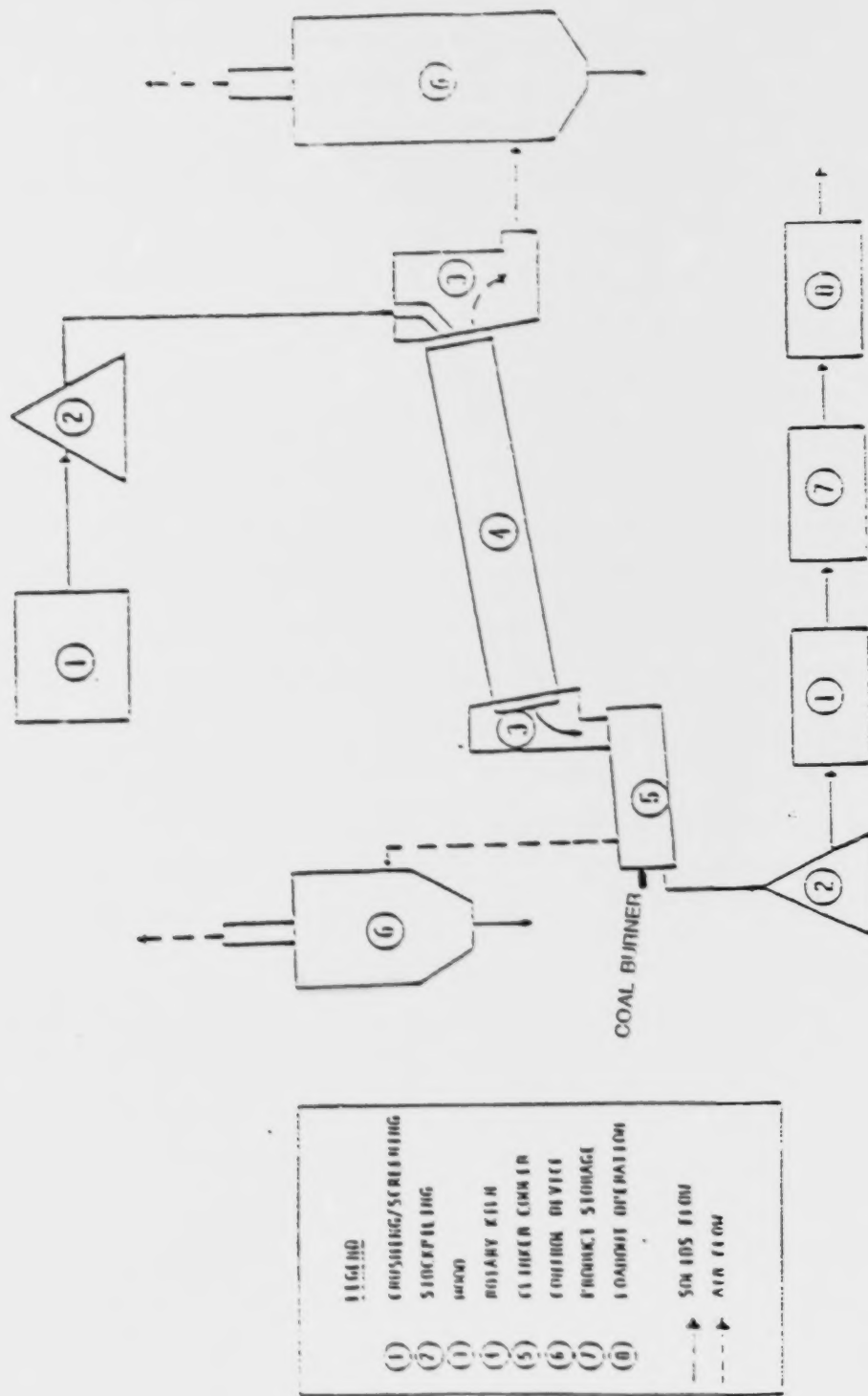


FIGURE 3.4 - SCHEMATIC DIAGRAM OF A TYPICAL LIGHTWEIGHT AGGREGATE PLANT (REFERENCE 2)

cooling down the kiln. Visual observation of the kiln contents is the usual method of determining that the desired lightweight aggregate is being produced and not some form of the undesirable alternatives. The temperature of the clinker at the discharge is about 1600°F. The kiln off-gas, at a temperature of about 500°F, is vented through an air pollution control system to an exhaust stack.

Combustion

Kilns are typically 7 to 10 feet in diameter and 120 to 300 feet long and are refractory lined. Kiln rotation rates are 60-100 revolutions per hour with solid material traveling down the sloped (3 to 4°) kiln and hot gases in the opposite direction. Production rates range from 250 to 1000 tons per day (Reference 8). Fuels include natural gas, oil, and coal, with a trend toward increasing coal use. The energy required per ton of product is 3 to 6 x 10⁶ Btu/ton of lightweight aggregate.

The hot clinker material discharged from the rotary kiln is air-cooled to near ambient temperature. For this cooling operation, moving grate or rotary cooler equipment is commonly used. Some of the off-gas from the cooling operation is used for combustion air for the kiln burner while the balance is passed through an air pollution control system to an atmospheric vent stack. Because of the necessity of balancing the gas flow through both the cooler and the kiln, this off-gas is not passed through the control device installed on the kiln. If necessary, the cooler section will have its own control devices.

Emissions and Air Pollution Control

Table 3.4 summarizes the air pollution systems used in the domestic production of lightweight aggregate by the rotary kiln process (Reference 9). These data show that for 36 producing facilities in 1979, 25 plants (or 69%) used wet scrubbers to remove particles from the kiln off-gas stream before venting to the atmosphere. Six plants (approximately 17%) used fabric filters for the same pollution source and one plant used an ESP. As discussed above, only 30 plants were operating in 1983. No update of the data presented in Table 3.4 was found.

BLAST FURNACES (STEEL INDUSTRY)

General

A blast furnace is essentially a large countercurrent flow chemical reactor for reducing iron ore to pig iron for use in the iron and steel industry. Because of its function, it is the keystone of this industry. Raw materials and energy are supplied to the furnace and molten iron and slag are withdrawn by several major auxiliary equipment items which are typically arranged as depicted in Figure 3.5. The major raw materials include iron ore, coke and limestone. Energy is provided by combusting the coke supplemented usually with one of a number of fuels.

TABLE 3.4

AIR POLLUTION DEVICES INSTALLED AT LIGHTWEIGHT AGGREGATE PLANTS

Number of Plants Employing Control Systems	System Type	Percent of Total Plants
25	Wet Scrubber	69.4
6	Fabric Filter	16.7
1	Electrostatic Precipitator	2.8
<u>4</u>	No Data	<u>11.1</u>
36		100.0

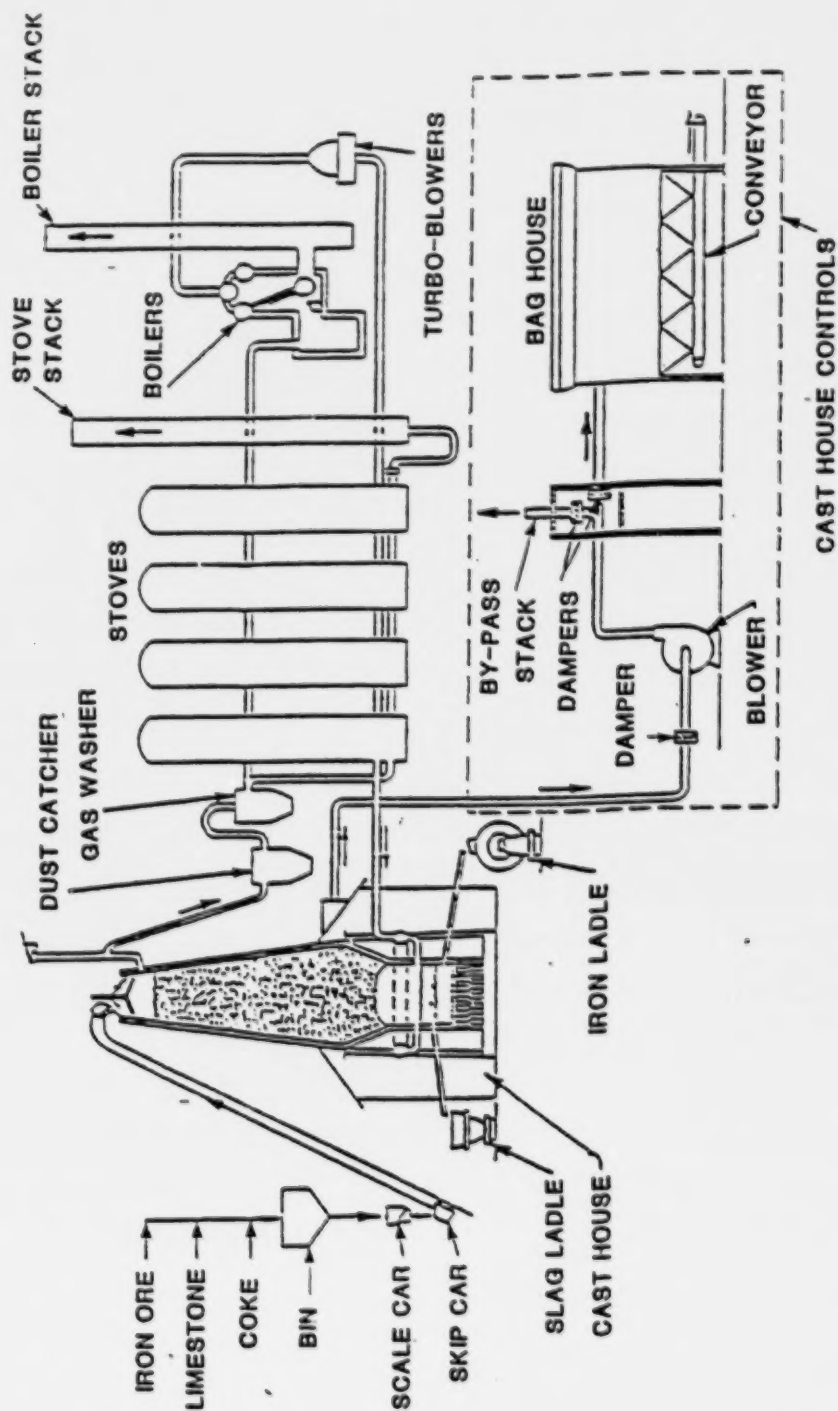


FIGURE 3.5 BLAST FURNACE WITH AUXILIARY EQUIPMENT

REFERENCE 10

The principal reactions are between coke, air, and iron ore. Part of the coke combines with the oxygen in the air to produce the principal iron ore reducing agent, carbon monoxide. This reaction generates the required heat for the process. Another part of the coke reacts with the iron oxide in the ore to release free iron which melts, drips to the bottom of the furnace and collects in the hearth. A final portion of the carbon, approximately 4%, dissolves in the iron (Reference 10).

Other important reactions are involved in the removal of impurities from raw materials. Limestone is the principal fluxing material used for removing the impurities. It is calcined by the heat of the hot blast air in the furnace. The resulting CaO reacts with the impurities in the ore, principally sulfur, to form a slag, which in the molten form, descends to the hearth. Because the slag is less dense than the molten iron collected in the hearth, it floats in a separate layer on it.

When the furnace hearth becomes full of the molten iron and slag, usually every three to four hours (Reference 10), tap holes in the furnace wall are unplugged. The molten iron is drained through an opening, known as an iron notch, into a ladle for transport to the steelmaking facility or to a pig casting machine. The molten slag is removed through an opening, known as a cinder notch, located at a higher elevation than the iron notch. The slag discharged from the blast furnace is either collected in a slag ladle for transport to disposal or into a slag pit. Once solidified, the slag may be used as aggregate for concrete, road bed, etc.

The energy needed for the iron smelting process is supplied by the combustion of the coke supplemented by some other fuel as previously mentioned. Thermal economy demands that the combustion air be preheated to about 1900°F to 2000°F before it is blown into the furnace. Preheating is achieved by passing the combustion air through a vertical regenerator chamber known as a stove, which is heated by firing some of the gases generated by the blast furnace reducing reactions. Each furnace utilizes three to four stoves, one of which is heating the blast air while the others are being heated by burning the blast furnace gas. Because this gas is laden with dust, it is passed through a high efficiency dust removal system before being combusted in the stoves.

Process Description

A blast furnace is a vertical, cylindrical structure that is composed of four major sections: the top, the inwall, the bosh, and the hearth (see Figure 3.6). The top section receives the charged burden and collects the blast furnace gases before it exits the furnace and serves to dry and preheat the solid raw materials, commonly termed "burden". The burden is charged into the top section of the furnace through a system of two or three pressure sealing bells. This multiple-bell charging system allows the solids to be charged while preventing the escape of the blast furnace gases. A blast furnace is normally operated under high positive pressure (approximately 100 inches water at the top of the furnace) to obtain increased production (Reference 10). The walls of the top section are vertical and lined on the inside with steel wear plates to protect the lining from abrasion from the charge. Hot gases rising from the lower sections of the furnace dry and preheat the raw solid materials in this top section. These gases accumulate in the top

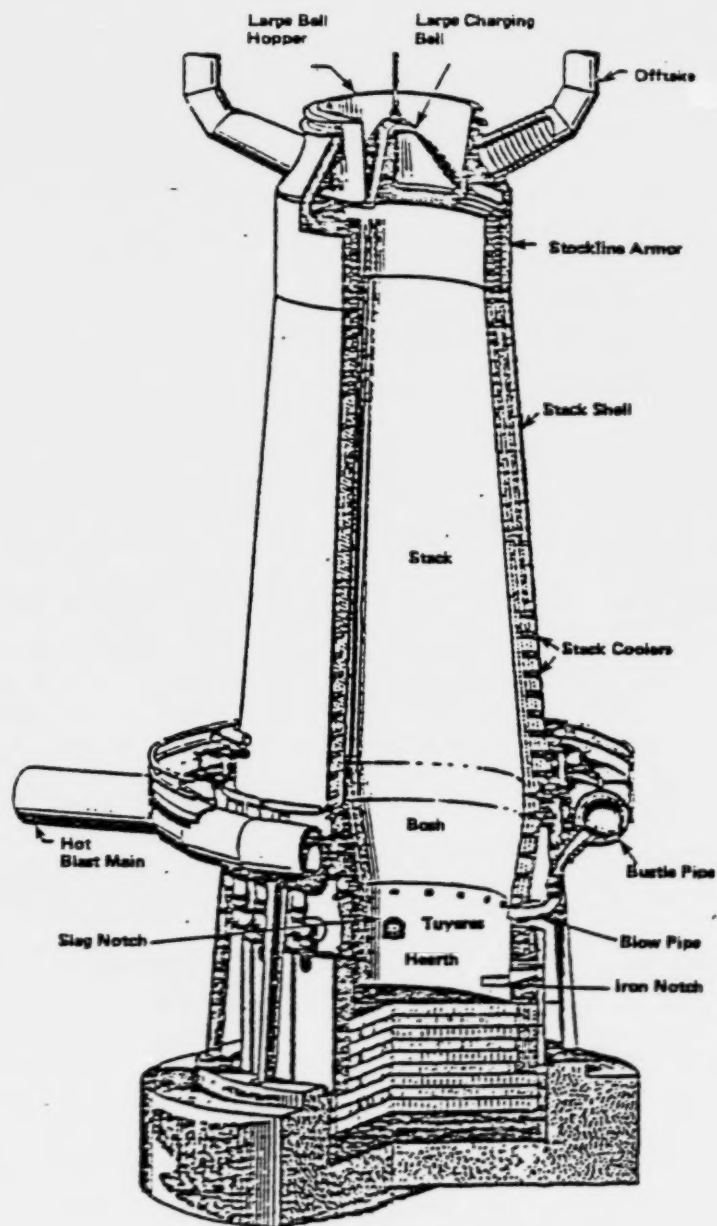


FIGURE 3.6 TYPICAL BLAST FURNACE

REFERENCE 10

section and are discharged through several off-takes (ducts) to a dust cleaning system.

The main section of the furnace where the iron oxides are reduced to free iron is known as the inwall. It is tapered outward from the top to the bosh section below. It is lined with firebrick and the lower portion is water cooled by means of steel cooling plates which are embedded in refractory.

The coke combustion occurs in the bosh section which extends from the widest part of the inwall down to just about the level where the combustion air is injected through nozzles known as tuyeres. This section is usually lined with super-duty firebrick and is cooled by cooling plates.

The bottom section of the furnace is known as the hearth. It extends from just above the tuyere level to the very bottom of the furnace. The molten iron and slag formed in the upper sections of the furnace accumulate in the hearth before they are periodically removed. Heated combustion air enters the furnace in the hearth section of the furnace through the tuyeres.

Encircling the bosh section is a manifold for the hot combustion air called a bustle pipe. It directs the hot blast air from the furnace stoves to the tuyeres located around the perimeter of the hearth. Depending on the size of the furnace, there may be from approximately 10 to 40 tuyeres. Air is conducted from the bustle pipe to the tuyeres through a series of connections as depicted in Figure 3.7 (Reference 11). The connections are designed for easy removal of the tuyere nozzle and blow pipe. The blow pipe is important in the consideration of burning hazardous wastes because it the most strategic point for injecting these wastes into the furnace as explained later.

Blast furnaces vary considerably in size and capacity. The 90 furnaces in operation in the United States in 1983 ranged from approximately 90 to 150 feet high with hearth diameters of 17 to 45 feet (Reference 12). Table 3.5 shows the location of these blast furnaces. The average height and hearth diameter are 102 and 27 feet, respectively. Capacities typical range from 550 to 7,000 tons of iron product per day although units of 10,000 tons per day capacity have been built (Reference 13). The U.S. 1983 production of iron produced in blast furnaces was 48.7 million tons (Reference 14).

Combustion

The smelting of iron ore in a blast furnace is highly endothermic. Thermal energy for the process is provided primarily by the combustion of the coke in a relatively small zone above the tuyeres. Here the carbon in the coke reacts with the oxygen in the hot blast air to form carbon monoxide and carbon dioxide. At the approximately 2,800°F temperature in the combustion zone (Reference 11), the carbon dioxide is unstable and reacts with the coke to form additional carbon monoxide. The reaction of the coke with oxygen is exothermic, but the reaction of coke and carbon dioxide is endothermic. However, the overall conversion of the coke to carbon monoxide is quite exothermic.

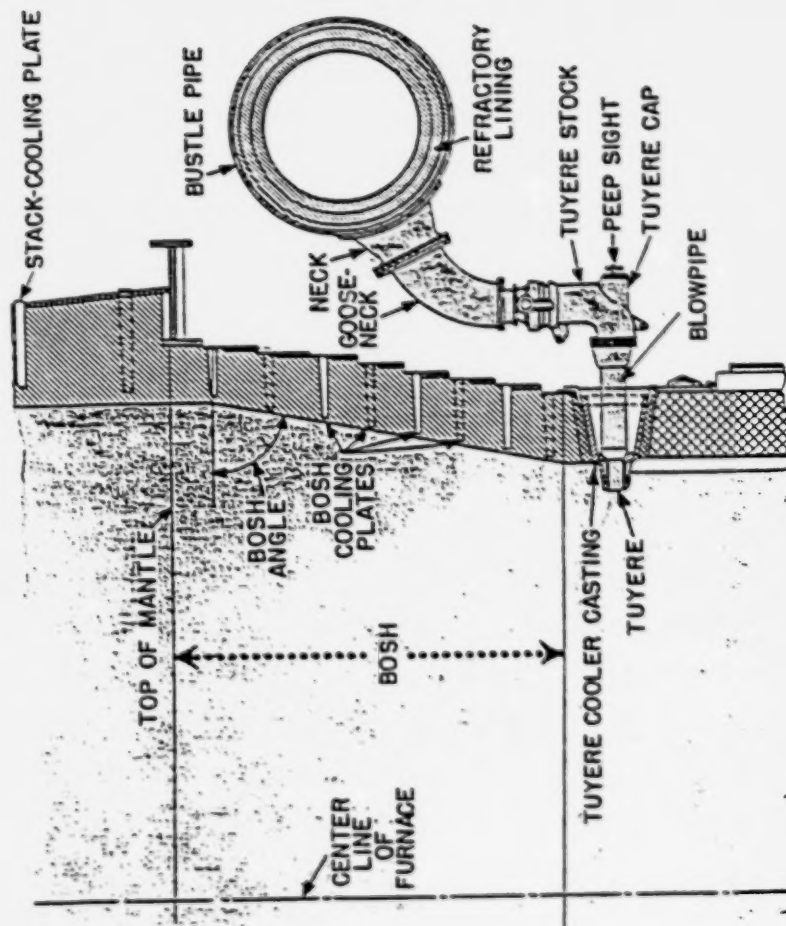


FIGURE 3.7 SECTION THROUGH TUYERE OF BLAST FURNACE

REFERENCE 11

TABLE 3.5

BLAST FURNACES IN THE UNITED STATES

Company	Location	Number of Blast Furnaces
Armco, Inc.	Ashland, Kentucky	2
	Hamilton, Ohio	2
	Middletown, Ohio	1
Bethlehem Steel Corp.	Bethlehem, Pennsylvania	4
	Burns Harbor, Indiana	2
	Sparrows Point, Maryland	4
Inland Steel	East Chicago, Indiana	9
Interlake	South Chicago, Indiana	2
Lone Star Steel	Lone Star, Texas	1
LTV Steel	Aliquippa, Pennsylvania	3
	Warren/Niles, Ohio	1
	Youngstown, Ohio	2
	Cleveland, Ohio	6
	East Chicago, Indiana	3
	Gadsden, Alabama	2
	South Chicago, Illinois	1
McLouth Steel Products Corp.	Trenton, Michigan	1
National Steel Corp.	Granite City, Illinois	2
	Ecorse, Michigan	3
Rouge Steel Co.	Dearborn, Michigan	3
Sharon Steel Corp.	Sharon, Pennsylvania	2
Shenango, Inc.	Neville Island, Pennsylvania	2
United States Steel Corp.	Braddock, Pennsylvania	3
	Fairfield, Alabama	2
	Fairless Hills, Pennsylvania	3
	Lorain, Ohio	4
	Gary, Indiana	5
	Geneva, Utah	3
	Weirton, West Virginia	4
Wheeling-Pittsburg	Steubenville, Ohio	5
	Wheeling, West Virginia	3
Total		90

Source: Reference 12.

Coke is essentially the sole primary raw material for providing the heat and reducing agent for the modern tall blast furnace because of its physical and chemical properties. Physical characteristics of greatest importance is its hardness and compressive strength. These properties give coke the necessary high abrasion resistance and the ability to support the burden and allow the hot furnace gases to rise uniformly through the burden. The chemical property of importance is its ability to be effectively utilized in the smelting and combustion reactions.

Other primary fuels, such as charcoal and coal, were used in the developing stages of iron smelting when much shorter furnaces were employed. Some of these materials are still used as the principal fuel in the few low-shaft furnaces used for small-scale iron production.

Since the late 1950's, the injection of an auxiliary fuel into the blast furnace has become a common practice to reduce coke consumption and to improve control of the furnace temperature. The economy of iron production is strongly dependent upon the amount of coke required to produce a ton of iron product. The amount of coke required to produce a ton of pig iron has dropped from 1,760 (in 1955) to 1,260 (in 1969) pounds of coke per ton of pig iron, and in some furnaces the rate is below 1,000 pounds per ton (Reference 11). Fuel injection is one of the effective methods being employed for reducing the coke requirement.

Several fuels have been used for this purpose including natural gas, coke oven gas, oil, tar, pitch, pulverized coal, coal-oil slurries, and recently, waste organic solvents. Although the method of injection differs with the type of fuel, the injection location is generally at the blow pipe 6 inches to 1-1/2 feet or more back from the nose of the tuyere (Reference 11). When a liquid fuel or coal injection is used, a lance pipe is generally inserted diagonally through the side of the blow pipe by means of an airtight connection to conduct the fuel into the air blast. Alternatively, it may be injected through a louver in the blow pipe face plate beside the peep sight. Gases may be injected through specially designed openings in the tuyere nose or through a lance in the blow pipe. The lance for liquid fuels are generally made by the steel company's maintenance department and may employ an atomizing nozzle.

The injection of hydrogen bearing fuels into the blast furnace brings about considerable change in the composition and properties of the ascending furnace gases (Reference 11). The fuel injection is more complex than the simple replacing coke carbon by fuel carbon units. For this reason, the injection rates are generally limited. The limiting quantity varies with the fuel. For natural gas, the injection rate is typically limited to between 1,000 and 2,000 standard cubic feet per ton of metal (40 to 90 pounds per ton) and for oil between 50 to 150 pounds per ton (Reference 11).

Firing of a Hazardous Waste Derived Fuel (HWDF) requires provisions for injecting depending on the compatibility of the HWDF with the other auxiliary fuel. Modifications to the auxiliary fuel injection system should not be required if the HWDF is of similar properties, e.g., the system used for injecting fuel oil could also be used for injecting many liquid organic solvents.

Some modifications to the auxiliary fuel injection configuration may be required if the HWDF only supplements another auxiliary fuel. The HWDF could be blended with the fuel prior to injection or a separate fuel injection system be provided. Blending the waste with the other auxiliary fuel requires changes in the fuel handling system but none in the burner gun. This approach requires the HWDF be compatible with the other fuel being injected. Also, special control provisions are needed to switch from the blended fuel to one containing no hazardous waste in the event of a process upset.

Emissions and Air Pollution Control

Approximately three tons of gases are evolved for every ton of iron produced (Reference 15). The volume and composition of these gases can vary significantly with how the furnace is operated and the amount of preparation of the burden. The amount of dust leaving the furnace can range from 0.01 to 0.15 unit weights of dust per unit of iron produced with concentrations as high as 69 grains per standard cubic foot of gas (Reference 10). Other major constituents include carbon monoxide (23 to 40 volume percent), carbon dioxide (12 to 16 volume percent), hydrogen (2 to 6 volume percent), moisture (2 to 12 volume percent) and nitrogen (30 to 67 volume percent) (References 10, 16-18). Gases leave the furnace at temperatures of 350 to 540°F (Reference 16).

Because of its high carbon monoxide content, the blast furnace gas has a heating value of about 75 to 90 Btu per cubic foot, and is used as a fuel within the steel plant. A survey (Reference 14) of 18 blast furnace facilities found that 41.3% is burned in the stoves, 52.2% in boilers, 2.0% for underfiring coke ovens, 3.0% is flared with the remaining gas going to soaking pits and combustion engines.

Before the blast furnace gas can be used as a fuel as described above, it must be cleaned of particulate matter. Otherwise, particulate matter would tend to deposit in the combustion zones of these units causing premature outages and failures. Since these units are necessary to the operation of the iron making process and since they require a high investment of capital, cleaning the blast furnace gas is deemed an essential part of the iron making process, aside from any environmental consideration.

To obtain the high level of particulate matter removal needed to use the blast furnace gas as a fuel in the steel making plant, two or three stages of gas cleaning are employed. The first stage consists of either a settling chamber or a dry cyclone to remove the large particles. The vast majority of steelmakers employ two stages of cleaning with a venturi scrubber providing the secondary cleaning (Reference 12). A few two stage dust removal systems use an electrostatic precipitator as the second stage device. Where three stage systems are used, the second stage is generally a wet scrubber and either a venturi or an electrostatic precipitator provides the third stage. Any of these combinations are capable of providing a clean fuel of less than (0.01 grains per cubic feet) for use in the steel plant (Reference 1).

Fugitive emissions to the atmosphere occur during the production of iron during hot metal transfer operations in the cast house and when there is a blast furnace "slip". A "slip" is a process upset where a stratum of the material charged to a blast furnace does not settle with the material below it, thus leaving a gas filled space between the two portions of the charge. When this unsettled stratum of charge collapses, the displaced gas may cause the top gas pressure to increase above the safety limit, thus opening a counter weighted bleeder valve to the atmosphere.

There are no federal New Source Performance Standards (NSPS) for control of air emissions from blast furnaces in the iron and steel industry. Since the blast furnace exhaust gases are not directly vented to the atmosphere, control of emissions while burning a HWDF would be accomplished by the internal process control requirements previously described.

Compatible Wastes

There are several materials that are known to adversely affect the pig iron quality and/or the blast furnace equipment. Sulfur and titanium bearing compounds affect the quality of the pig iron. The molten iron reacts with sulfur to form ferric sulfides which alter the properties of the pig iron. Titanium forms compounds with carbon and nitrogen which are soluble at high temperatures but have limited solubility at normal temperatures. These compounds come out of solution and make the metal very viscous when it flows from the furnace and starts to cool. Consequently, the iron freezes in the runners and causes handling problems.

Alkali metals have several major adverse effects. They attack the refractory in the furnace as well as the stoves used to heat the blast air and are an important contributor to the formation of furnace wall deposits and furnace bridging which leads to slips (Reference 11). Alkali metals also reduce the iron production and increase the coke rate (Reference 11).

Halides, in high concentrations, attack the furnace and stove refractory and contribute to the formation of slips. For these reasons, the rate of halogenated compound firing as auxiliary fuel must be limited.

Lead and zinc are two other very troublesome metals. Lead is readily reduced in the blast furnace but is almost insoluble in the hot metal. Amounts of lead greater than can be volatilized as fume may collect beneath the molten iron in the hearth with disastrous effect on the lining (Reference 11). Zinc vaporizes and travels upward with the furnace gas. Some of the vapors condense in the upper part of the furnace lining next to the shell. Unless the top temperature is high enough, this material will accumulate in the furnace top and split the lining or the shell.

Because the blast furnace is the keystone of a steel plant, upsets in the operation caused by the firing of waste would be extremely costly. Steelmakers are, therefore, expected to be very cautious to limit the content of the materials described above in hazardous wastes fired as auxiliary fuel. An upset in furnace operation resulting from firing wastes could quickly offset the fuel savings realized from this practice.

In addition, there is concern of the release of undestroyed toxic wastes with fugitive emissions. The casting and flushing of slag are more common sources of fugitive emissions. Because gases escape from the tap holes, injection would have to be stopped during these operations. Fugitive emissions can also occur at the top of the furnace as a result of leaking seals in the charging system or during slips. It is unlikely that high concentration of toxics would escape with the off-gases. As mentioned in Section 3, the EPA Burner Survey showed three blast furnaces as burning a hazardous waste in 1982. These facilities reported the following:

- o Facility 1 - 208,000 gallons/year of tar sludge and coke oven by-products.
- o Facility 2 - Plant 1 - 5,861,729 gallons/year of recovered solvents still bottoms and other wastes.
Plant 2 - 3,907,819 gallons/year of similar wastes.

REVERBERATORY FURNACES (OPEN HEARTH FURNACE - STEEL INDUSTRY)

General

A reverberatory furnace delivers heat to the furnace charge by directing a long flame above the charge. The flame not only radiates heat directly to the charge supported on the furnace hearth but also heats the refractory in the furnace roof and walls which radiate and reflect thermal energy to the charge. The term reverberatory is derived from the fact that radiant energy provided by the flame is reflected from the roof and walls onto the charge.

Reverberatory furnaces are found in numerous applications. The major use has been in the production of steel in the open hearth furnaces which is a reverberatory furnace. They are also widely used as remelt furnaces in the aluminum industry. Reverberatory furnaces are used for smelting copper, lead, zinc, and tin. The detailed analysis of reverberatory furnaces will be limited to open hearth furnaces. There were 44 open hearth furnaces in operation in 1984 (Reference 12) as indicated by Table 3.6. These varied in capacity from 262 to 420 tons per heat. No new furnaces are being constructed and old ones are being phased out because of higher cost of production per ton and lower productivity compared with basic oxygen furnaces. The open hearth furnace accounted for only slightly over 7% (5,951,000 tons) of the total 1983 United States steel production (Reference 19).

Process Description

The open hearth furnace is used to produce steel from a mixture of scrap and hot metal in varying proportions. Production of steel in the furnace is a batch operation and consists primarily of the melting the solid materials, oxidizing the impurities, reducing the iron oxides to iron and adding the desired alloying constituents. The major impurities include carbon, silicon, manganese, phosphorous and sulfur. Major raw materials include the pig iron, scrap iron and a fluxing agent (usually limestone). Iron ore may also be added to provide low cost iron units and oxygen for controlling the silicon and carbon content of the steel. Oxygen lancing is used to accelerate the refining period.

TABLE 3.6

UNITED STATES OPEN HEARTH FURNACES OPERATING IN 1984

Company	Location	Number of Furnaces	Capacity each Furnace (tons/heat)	Air Pollution Control System
Armco, Inc.	Middletown, Ohio	6	310	venturi each furnace
Bethlehem Steel Corp.	Sparrows Point, Maryland	7	420	6 ESPs & 2 ven- turis connected in parallel
Inland Steel Co.	East Chicago, Indiana	7	335	5 ESPs connected in parallel
Lone Star Steel Co.	Lone Star, Texas	5	262	3 steam-hydro scrubbers connected in parallel
United States Steel Corp.	Fairless Hills, Pennsylvania	9	400	9 ESPs connected in parallel followed by 1 secondary ESP
	Geneva, Utah	10	340	8 ESPs/venturi combinations connected in parallel

Source: Reference 12.

A typical "heat" (a tap-to-tap cycle) is completed in the following sequence:

- o Charging the scrap into the furnace through hydraulically or electrically controlled charging doors.
- o Adding molten metal through the charging doors by pouring the metal from a ladle into a receiving trough.
- o Injecting oxygen into the molten metal and analyzing the metal until the impurities are removed.
- o Adding the desired alloying agents.
- o Tapping the slag and molten steel.

Oxidizing of the impurities takes place by the reaction with oxygen from the ore, from the decomposition products of the limestone and from the jet of pure oxygen injected into the bath. The removed impurities accumulate in a continuous layer on the surface of the liquid metal.

When the molten metal bath is determined to be the grade of steel required, the contents of the furnace are drained through the tap hole into steel ladles. The slag, which floats on the top of the metal, overflows into a slag pot when the ladle has been filled with steel. Alloying, re-carbonizing and deoxidizing agents are added to the molten metal before the slag starts flowing.

The tap-to-tap time of the furnace not using oxygen lancing is typically 8 to 10 hours compared to 4 to 5 hours with lancing (Reference 20).

Figure 3.8 shows the typical open hearth furnace design used in the steel industry. The open hearth furnace is a large reverberatory furnace consisting of two major sections: (1) the melting chamber, and (2) the regenerator (checker) chambers. The melting chamber is where the actual steel production takes place. This chamber is a large shallow basin (hearth) enclosed by refractory bricked walls and roof. Provisions are contained in the walls for raw material charging and for molten metal and slag removal. Heat is supplied from burners at each end of the chamber. The flame travels the length of the chamber above the charge which rests on the hearth. The front wall which contains charging doors are straight but the back wall slopes to facilitate the flow of metal out of the furnace during tapping. Near the bottom of the back wall is a tapping spout for removing the molten slag and metal. The furnace is reverberatory in that the hearth is heated both directly by the burner flame and by the radiant heat from the furnace roof and walls.

Raw materials are charged through the charging doors by use of a charging machine, a large vehicle which moves on tracks in front of the furnace. Solid materials are placed in charging boxes which are positioned within the furnace by the charging machine. When the box is inside the furnace, it is rotated about its axis, dumping its contents into the furnace. To provide distribution over the hearth, there are at least five doors for charging. Molten metal is charged through the charging doors by pouring it from a ladle into the receiving trough.

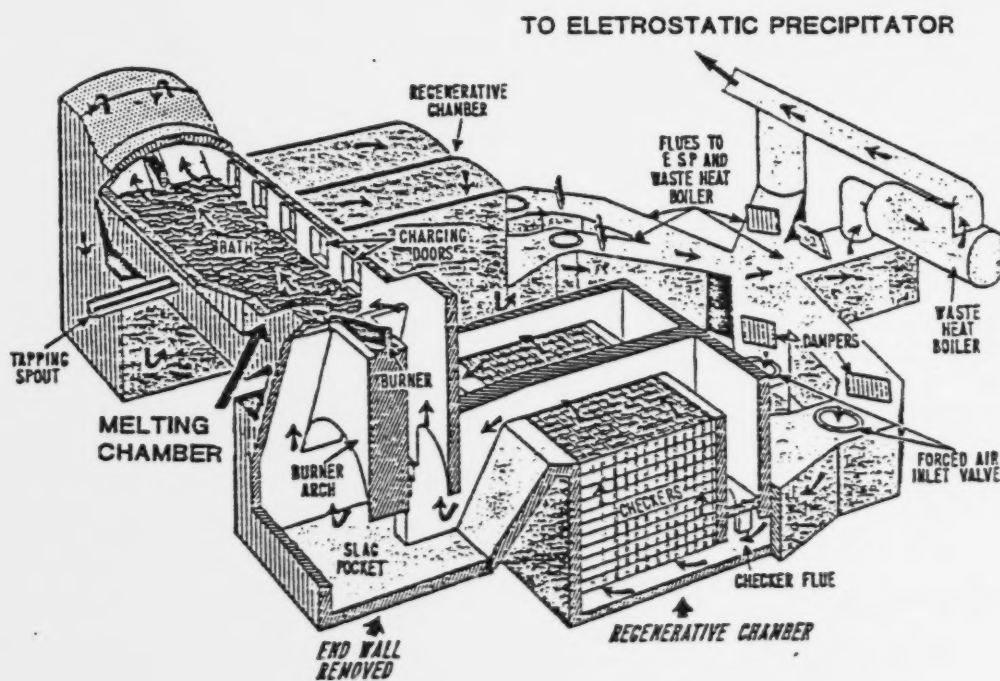


FIGURE 3.8 OPEN HEARTH FURNACE

REFERENCE 20

The open hearth furnace is regenerative, i.e., it uses the heat removed from the exhaust gases to preheat the combustion air. A regenerator chamber is located at each end of the furnace for this purpose. Hot gases from the furnace at temperatures of 3000°F (Reference 23) are discharged through the uptakes or ports at the ends of furnace, then through a slag pocket to remove heavy particles of slag carried by the gases, and finally through the regenerator chambers. Flow of combustion air and exhaust gases is in a single direction at any given time, with combustion air being preheated in one regenerator chamber while the furnace exhaust gases are being cooled in the other. The flow is reversed periodically to permit the chambers to alternately be heated and cooled and to maintain a reasonably uniform combustion air temperature.

The furnace melting chamber is quite large providing gas retention times of 2 to 3 seconds (Reference 23). Dimensions vary widely with the furnace capacity. The smaller furnaces in operation today typically have inside bath dimensions of about 20 feet in width by 50 feet in length and height from hearth to roof of 11 feet (Reference 13). The larger furnaces are typically about 30 feet in width and 75 feet in length (Reference 13). The long residence time coupled with the high temperatures (3500°F flame temperature and exit gas temperature of 2700°F) operation makes these furnaces suitable for destroying hazardous wastes (Reference 23).

Combustion

Heat for the production of steel in the open hearth is provided by the combustion of a combination of the following fuels: natural gas, oil and tar and pitch. All furnaces are equipped to burn both a liquid and gaseous fuel. The heat input requirement is 3 to 4 million Btu per ton when molten iron is charged compared to 4.2 to 5 million Btu per ton with all cold scrap (Reference 20).

The fuel is fired through two long (approximately five feet) combination burners (i.e., capable of firing both gaseous and liquid fuels) located at opposite ends of the melting chamber. They are designed to throw a very long flame across the furnace hearth to generate the high temperature needed for steel production. Since the liquid fuel burned may often be quite viscous, some type of atomization is generally provided. Fuel firing alternates between burners. Alternating of firing between burners is synchronized with the furnace gas flow reversal such that the burners are always firing in the direction of flow. Switching of the burner firing is automatically controlled.

Firing of the burners begins after the scrap is charged and is continuous until the analysis of the metal indicates that the desired carbon-temperature relationship is achieved during oxygen lancing. Thereafter, the burners are fired intermittently to maintain the proper temperature. The firing is controlled manually; monitoring of combustion gas constituents to ensure good combustion is not practiced.

Air for combustion is delivered from blowers through the preheating regenerator chambers into the ports in which the burners are mounted. A set of dampers is provided in the ductwork to allow for the reversal of gas flow. Air enters the ductwork through a forced air inlet valve; passes through the checkers, rises through the uptakes at the end of the furnace and passes by

the burner where it combines with the injected fuel. The combustion products pass over the bath, go down the uptakes on the other side of the furnace, pass through the checkers giving up heat to the brick, and exit through the waste heat boiler to control device followed by the stack.

Emissions and Air Pollution Control

Both primary and fugitive air emissions result from the operation of an open hearth. Primary emissions refer to those emissions leaving the furnace through the regeneration chambers. Fugitive emissions from the furnace result from ancillary operations such as furnace charging, tapping and flame reversal.

Particulate matter is the major pollutant found in the primary emissions. These are generated by the refining process and by combustion of the furnace fuels as well as the oils and grease on the metal scrap. The greater portion of the particulate matter is formed by the refining process and is emitted from the furnace as metallic oxides (Reference 24). These particulate emissions consist of very small particles with nearly 50% being smaller than 5 micrometers (Reference 25). Uncontrolled particulate emissions range from 9.4 to 22 pounds per ton of steel produced (Reference 16). The concentration of the particulates in the gas stream varies over a wide range during a heat with 0.7 grains per standard cubic foot being an average value (Reference 24).

The small particle size of the particulate emissions necessitates the use of a high efficiency collection device. As Table 3.5 indicates, electrostatic precipitators, venturi scrubbers, and combinations of these devices are used to control particulate emissions from open hearth furnaces. Most of the air pollution control systems currently in operation have a common manifold to distribute the gas flow from several furnaces evenly among the different control devices. Electrostatic precipitators have demonstrated over 98% collection efficiency (Reference 24). Outlet dust loadings of less than 0.01 grains per standard cubic foot can be achieved with a venturi scrubber operating with a pressure drop of over 40 inches of water column (Reference 25). There is no NSPS for open hearth furnaces but there are standards for other types of reverberatory furnaces. Primary lead, brass and bronze, and copper and secondary lead reverberatory furnaces have NSPS for particulate emissions of 0.022 gr/dscf.

The gases leave the furnace melting chamber at about 3000°F and are cooled to about 1300°F as it passes through the regenerative chamber (Reference 20). Where an electrostatic precipitator is to be used, the gases are passed through a waste heat boiler to cool them to a temperature of 500 to 600°F (Reference 20).

Fugitive emissions from the furnace occur during molten metal transfer, furnace charging, tapping, and flame reversal as previously mentioned. Although the potential environmental impact of these emissions are a major concern of burning hazardous waste, insufficient information was found to permit their quantification.

Two methods of injecting liquid HWDF are possible. The method requiring the least modification to the fuel handling and firing system would entail blending the waste with the other liquid fuels being fired. This required that the waste be compatible with other liquid fuels being used. Provisions

for switching from waste containing fuel to a conventional fuel during operations generating significant fugitive emissions (furnace charging and flame reversal) is deemed necessary.

Another plausible liquid waste injection method is to provide a separate fuel handling system and burner gun for the HWDF. A waste burner guns could be inserted into the furnace adjacent to existing guns (one at each end of the furnace). This approach allows independent control of the waste and other fuel firing. Separate fuel trains and burners are needed to fire wastes that are incompatible with conventional fuels.

Compatible Wastes

Burning HWDF in an open hearth furnace could adversely affect both the furnace and the steel quality if certain materials are not avoided. As with blast furnace, halogenated solvents and waste containing alkali metals, lead and tin should be burned in limited quantities because they could reduce the life of the refractory lining. Damage to the refractory which results in loss of operating time could easily offset the fuel savings realized from burning hazardous wastes. The burning of halogenated solvents would also generate hydrochloric acid which could cause serious corrosion of the air pollution control equipment.

The impact on steel quality would be from metals and inorganics in the waste as the organic compounds should essentially be destroyed at the high temperature of over 3000°F and residence time of 2 to 3 seconds (Reference 23) common for an open hearth furnace. These materials would be partitioned between the flue gases, slag, and molten metal. Because the flame is directed over the surface of the furnace hearth it is expected that little of these materials would be absorbed in the slag and metal. Those that were absorbed, however, should be distributed between the slag and metal as shown in Table 3.7 based on their ease of being oxidized relative to iron. Since steel properties are determined by its chemical composition, the burning of wastes containing those elements that would remain in the molten metal could potentially affect the steel quality. To avoid potential quality deterioration, wastes should be monitored for the presence of any such elements. Acceptable levels of these contaminants in the waste could be determined from fuel firing rates and metal composition specifications.

Burning hazardous wastes in an open hearth furnace can result in toxic materials as fugitive air emissions during the charging of raw materials through the doors and during flame reversal. Toxic materials not completely destroyed in the furnace could possibly enter the work space as fugitive emissions if hazardous waste is burned during these two operations. Fuel firing could be controlled, however, such that no waste is injected during these operations.

SPENT ACID RECOVERY PLANTS

General

Spent acid recovery furnaces are used to recover sulfuric acid from spent acid. The spent acid is sulfuric acid that has been used in a process

TABLE 3.7

DISTRIBUTION OF WASTE CONTAMINANT ELEMENTS BETWEEN SLAG AND METAL

Group I	- Elements almost completely taken up by slag: Si, Al, Ti, Zn, B, V, Ca, and Mg
Group II	- Elements distributed between slag and metal: Mn, P, S, and Cr
Group III	- Elements almost completely in solution in metal: Cu, Ni, Sn, Mo, Co, W, As, and Sb
Group IV	- Elements out of slag and metal: Zn, Cd, Pb, Ag, and Hg

Source: Reference 23.

and has become contaminated with water, organics, inorganics, etc. The acid drops in strength and is no longer usable for its intended purpose. In a recovery furnace, the spent acid is thermally decomposed at elevated temperatures into SO_2 , carbon monoxide, carbon dioxide and water vapor. Heat needed for the acid decomposition is supplied by burning fuels, typically natural gas, oil or a liquid or gaseous waste stream. The SO_2 generated in the recovery furnace ranges from 0-14% of the exhaust gases.

The conversion of the SO_2 to sulfuric acid follows the contact process as depicted in Figure 3.9. The SO_2 is oxidized to SO_3 in a multi-stage converter, typically having either three or four stages depending on the age of the plant and overall conversion efficiencies desired. The converting from SO_2 to SO_3 requires very controlled reaction temperatures for efficient conversion. The exhaust gases from the combustion chamber are cooled by a waste heat boiler prior to entering the converter. After passage through each stage of the converter, the gases are cooled by waste heat boilers or super heaters to maintain the desired temperature ranges. The stages of the converter each have a bed of solid vanadium pentoxide catalyst. The conversion rate drops as each stage of the converter has less SO_2 than the previous stage. Gases leaving the converter are again cooled and are introduced into the SO_3 absorber at about 450°F. If oleum is to be produced, the temperature is lowered further prior to entry into the oleum tower.

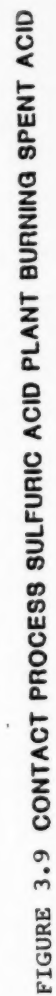
The SO_3 absorber is similar to the drying tower used for the combustion air. The SO_3 is absorbed in sulfuric acid which is continuously drawn-off as the product (Reference 26). The sulfur dioxide, sulfur trioxide and sulfuric acid remaining in the exhaust gases must be reduced to meet air pollution standards.

Since the enactment of these strict air pollution regulations in 1971, most new plants use a dual absorption process to lower the SO_2 emissions. The primary SO_3 absorber, known as the interpass absorber, is used as the gases leaving the third stage of the converter (Reference 27). The lower partial pressure of the SO_3 in the gases returning to the fourth stage increases the conversion rate and thereby lowers SO_2 emissions. The second SO_3 absorber is similar to the single absorber system but only absorbs SO_3 from the gases exiting the fourth stage of the converter.

Plants built prior to 1960, generally had three stage converters and overall conversion efficiencies of 95-96% (Reference 27). In modern four stage converter contact process plants the overall conversion efficiencies approach 99%+.

If oleum is produced at the acid recovery plant, the SO_3 leaving the converter is cooled further and is introduced to an absorption tower with 100% sulfuric acid.

The recovery furnace is usually a refractory lined cylindrical chamber equipped with a burner designed to spray the spent acid into the chamber without flame impingement on the walls. The design of these furnaces vary according to plant production capacity, but can be 40 feet long by 14 feet in diameter. The combustion air is sometimes dried before entry into the burner in order to prevent corrosion (acid mist formation) of the process equipment by the SO_3 combining with the water vapor to form sulfuric acid.



Since spent acid have contaminants such as water, organics and inorganics, they have the potential to corrode and plug the converter beds and other downstream equipment. It is reported that spent acid usually contains about 90% sulfuric acid, 4 to 5% water, and 5 to 6% hydrocarbons (Reference 28). The gases are cleaned and dried after the combustion/waste heat boiler by cyclones, electrostatic precipitators, scrubbers and/or gas cooling towers.

Combustion

The combustion of spent acid needs to be balanced to achieve the desired range of 8-14% SO_2 in the exhaust gases. This can be accomplished by fuel blending and combustion air adjustments. It is important to note that as the percentage of SO_2 drops, more exhaust gases need to be processed to create the same amount of acid (i.e., the plant exhaust gases with SO_2 at 4% in the combustion gases, will be 2.5 times that of SO_2 at 10% in the combustion gases) (Reference 27). Therefore, burning multiple fuels with/and without significant oxygen contents, makes the efficient combustion system very complex (Reference 29). The control of combustion relies on both temperatures and oxygen controls.

Emission and Air Pollution Control

The exhaust gases from a spent acid recovery furnace are processed to sulfuric acid before being emitted to the atmosphere. Therefore, furnace air emissions are referenced to the tail and of the entire acid plant. The major pollutants emitted from a spent acid plant are SO_2 and sulfuric acid mist. The quantity of SO_2 emitted is an inverse function of the sulfur conversion efficiency (SO oxidized to SO_3). This conversion is always incomplete, and is affected by the number of stages in the catalytic converter, the amount of catalyst used, temperature and pressure, and the concentrations of the reactants (SO_2 and O_2). Most single absorption plants have SO_2 conversion efficiencies ranging from 95-98%, which corresponds to 26 to 55 lb SO_2 emitted per ton of 100% H_2SO_4 produced. The EPA performance standard for new plants is 4 pounds SO_2 per ton of 100% H_2SO_4 produced which corresponds to a conversion efficiency of approximately 99.7%.

Nearly all the acid mist emitted from the spent acid plant is in the absorber exit gases. Acid mist is created when SO_3 combined with water vapor at a temperature below the dew point of SO_3 . Once formed within the process system, this mist is so stable that only a small quantity can be removed in the absorber. The water vapor is produced from the water in the spent acid and products of combustion of the fuels and hydrocarbon impurities in the spent acid. The strength of the acid produced also affects mist emissions. Plant producing higher concentrations produce greater quantities of finer more stable mist. Another greater quantities of finer more stable mist. Another factor affecting the mist emissions is the operating temperature of the SO_3 absorption tower. The optimum absorber operating temperature depends on the strength of the acid produced, throughout rates, inlet SO_3 concentrations. Typically, uncontrolled mist emission from a spent acid plant are 2.2 - 2.4 lb/ton acid produced (Reference 15). The EPA New Source Performance Standard is 0.15 pound per ton of acid produce which requires the application of control devices which are discussed later in this subsection.

When the spent acid and fuels burned in the recovery furnace, particulates, metal, and trace quantities of other pollutants may also be generated. The nature of these pollutants will depend on the type of fuel burned and the contaminants in the spent acid. Those pollutants that might plug or deactivate the converter catalyst must be removed before the furnace exhaust gases enter the converter. Thus air pollution control equipments are used in spent acid plants for pre-converter cleaning of the furnace exhaust gases (for particulate matter, metals, acid mist, etc.) and to the process exhaust gases prior to their being released to the atmosphere.

The pre-converter control can be:

- o Cyclones
- o Scrubbers
- o Electrostatic precipitators
- o Dryers

The emission controls used to meet NSPS are:

- o Sulfur dioxide control - dual absorption systems
 - sodium sulfate to sodium bisulfate scrubber
 - ammonia scrubber
- o Acid mist - electrostatic precipitator
 - mist eliminators
 - packed bed scrubbers
 - molecular sieves

The pre-converter controls are well suited to remove particulate matter, vaporized metals and hydrochloric acid emissions. All of these parameters would have operational significance if allowed to contaminate or plug the catalyst beds in the converter. While not all spent acid plants have all of these control devices, they must have adequate controls to prevent the expensive replacement of the catalyst.

Compatible Wastes

Hazardous waste burned in a spent acid recovery furnace as fuel must not adversely affect the acid production process equipment nor significantly alter the acid quality. The pre-converter cleaning devices can remove some hazardous waste contaminants that are not destroyed during the combustion, such as metals. However, because of the possibility that some portion of metals reaching the converter, operations are likely to restrict the level of metals in the waste fuel. They would also likely limit the amount of chlorinated wastes burned as fuel to avoid potential conversion problems in furnace and prevent adversely affecting the acid quality.

ASPHALT CONCRETE PLANTS

General

The bituminous concrete industry composes some 3,000-4,500 plants (References 30 and 31) across the United States. Their locations follow the general population and number of motor vehicles patterns since their product is used in population/vehicular related ways (i.e., roads, parking lots, driveways). Approximately 60% are stationary, 20% are mobile and 20% are somewhat transportable (Reference 30).

Bituminous concrete is a blended mixture of asphaltic cement (AC) and aggregate. The variations of the types of mixes are numerous with different asphaltic cement blends and different size and material combinations of aggregates (i.e., sand, rock, glass, asbestos, reprocessed asphalt pavement). The basic asphalt concrete manufacturing process is composed of:

- o Heating of the aggregate to drive off moisture and to facilitate mixture with the AC.
- o Screening of the aggregate.
- o Blending the aggregate with heated AC and mixing for a homogeneous product.

The actual plant designs are grouped into three subsets:

1. Batch-mix plants - separates the heating, screening, and blending production steps. Mixes one batch at a time; rotary dryer is started and stopped many times during the day. Represents 91% (Reference 30) of reported population.
2. Drum-mix plants - sizes the aggregate first, then combines the heating and blending steps. Usually runs for longer periods than a batch plant. Represents 2.5% (Reference 30) of the reported population but is the majority of all new plants.
3. Continuous-mix plants - similar to a batch-mix design but runs continuously for long periods based on constant demand. Represents 6.5% (Reference 30) of the reported population.

Process Description

For the context of burning HWDF's, the three subsets can be looked at as two groups; rotary dryer heating of the aggregate only (typical batch-mix plant) and rotary dryer heating and mixing the aggregate and AC (typical drum-mix plant). The continuous-mix design has the same combustion characteristics as the batch-mix design.

Figure 3.10 shows the schematic of a typical batch-mix plant. Aggregates are stored in outside piles of graded stone, sand, etc. The aggregates are fed into hoppers and then proportioned based on mix design, onto a conveyor belt to be fed into the rotary dryer. The aggregate is heated to about 300-350°F depending on mix requirements and then moved by a hot elevator to the screens for final sizing. The screens discharge into the pug mill where the hot aggregate is blended/mixed with the hot AC. After mixing the batch is

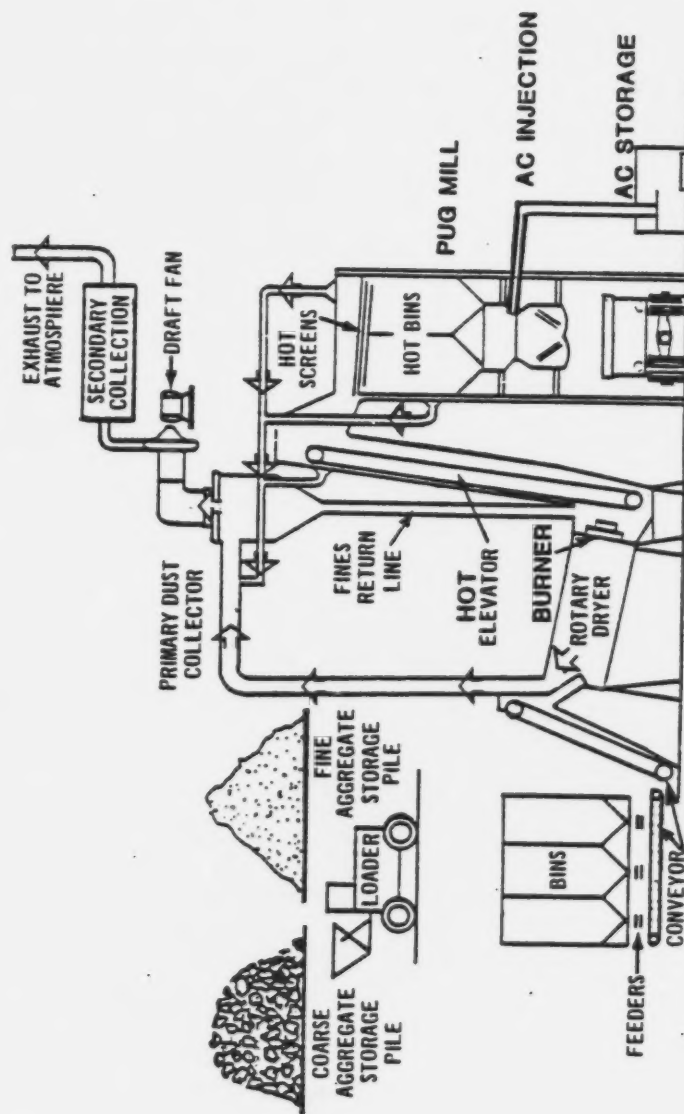


FIGURE 3.10 TYPICAL BATCH-MIX ASPHALT CONCRETE PLANT

REFERENCE 33

dropped into a dump truck for transporting to the job. Typically, several batches are required to fill a waiting truck and the system (i.e., conveyor, dryer, screens, pug mill) only operates on a per truck basis. Recent techniques have been applied to this operation that allow:

- o If the requirement for a certain mix is large (roadway paving/resurfacing), a heated storage silo is used to run the system for longer intervals and still allow for making other batch mixes.
- o Recycling of old asphalt pavement (RAP) is performed for economics of the roadway project and plant operations, as well as, reducing resurfacing pavement height and total pavement weight. The previous layer of asphalt concrete is removed and taken to the plant. It is crushed and sized as regular aggregate but stored separately. It is usually heated in an auxiliary dryer by the hot exhaust gases of the rotary dryer and combined with the hot aggregate for mixing with the AC in the pug mill.

The rotary dryers are designed to quickly heat the rock and drive off most of its surface and pore moisture that would interfere with the proper bonding with the AC.

The fuel being combusted (gas, oil, waste oil, HWDF) in the burner, is converted into heat and exhaust gases. Outside the actual flame, the gases contact the cold stone and quickly drop in temperature to about 350°F or less. The desired flame pattern is short and wide but without impingement on the sides of the dryer.

The typical drum-mix plant is shown in Figure 3.11. The aggregate is taken from the storage piles and is metered to the conveyor system for specific mix requirements. In the rotary dryer, the AC is sprayed onto the aggregate just outside of the flame zone (there are several variations to this AC injection method). The aggregate's moisture is driven off and the AC and aggregate are mixed by the rotating action of the drum dryer. No hot screens, hot elevators or pug mill are needed with the drum-mix plant. Due to its design of mixing in the dryer, efficient operation dictates more continuous production. Therefore, storage silos for the asphalt concrete are typically found at all drum-mix plants. Also, when RAP is used at these plants, it is introduced into the drum after the flame zone.

The drying/mixing zone of the drum-mix dryer, is similar to the dryer of a batch-mix plant but modified to prevent flame impingement on the AC. The exhaust gas temperatures of a drum-mix plant can be slightly higher than a batch-mix plant as are the temperatures of the asphalt concrete, to compensate for additional heat loss in storage.

Combustion

The combustion zones in both the batch-mix and drum-mix dryers are similar. The flame is short and releases sufficient heat to raise the aggregate to approximately 300°F. The burners use sufficient excess air to ensure proper combustion of the fuel and the desired aggregate temperature. Residence time at combustion conditions (>1,500°F) is short since the wet aggregate cools the exhaust gases very quickly. The burners operate efficiently

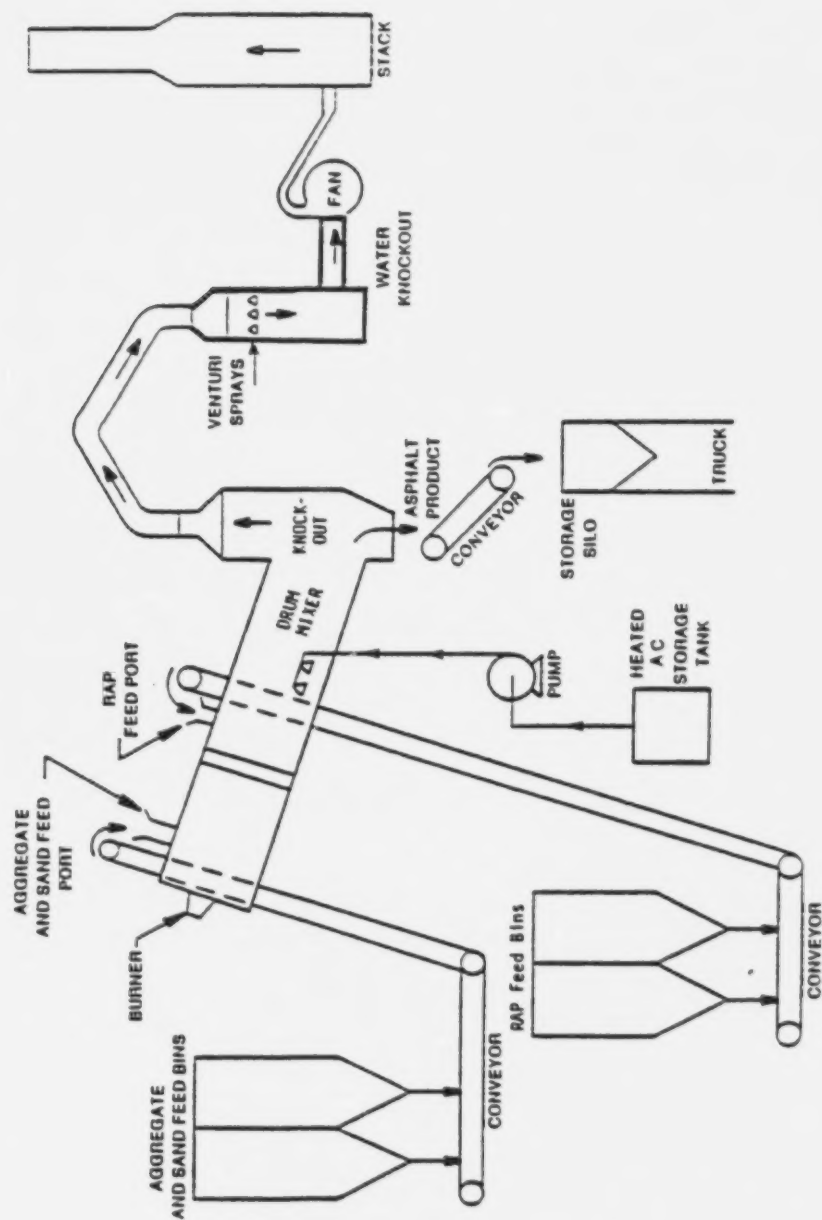


FIGURE 3.11 TYPICAL DRUM-MIX ASPHALT CONCRETE PLANT

for their normal fuels and have the potential to adequately combust some hazardous waste derived fuels.

Emissions and Air Pollution Controls

The New Source Performance Standards (NSPS) (Reference 32) for asphalt concrete plants are as follows:

1. Particulate emissions - ≤ 0.04 gr/dscf
2. Opacity - $\leq 20\%$

The standards apply to those plants constructed or modified as of June 11, 1973 (estimated to be 15% (Reference 30) of all asphalt concrete plants).

The compliance with the NSPS particulate emission standards typically requires the use of:

- o Baghouses
- o High energy scrubbers

The typical batch-mix plant can meet the NSPS standards by use of baghouse or scrubber, however, due to energy requirements, baghouses predominate (Reference 30).

The typical drum-mix plant uses a scrubber or baghouse (Reference 30) but the use of scrubbers is prevalent due to the potential blinding of baghouse filters with condensed hydrocarbons from the AC. It has been reported (Reference 33) that some drum-mix plants do not utilize any controls due to the inherent particulate control of the AC as it forms a collection surface for particles generated from the aggregate. However, the AC and the heat in the drum produces a hydrocarbon emission, that condenses and cause condensible particulate emissions and plume opacity (Reference 30).

Plants built prior to 1973 are principally the batch-mix or continuous-mix design and use air pollution controls such as:

- o Cyclones
- o Low energy scrubbers

The typical emission regulations for these older plants are process weight related and allow particulate emissions an order of magnitude greater than NSPS. The typical process weight derived allowable particulate emission rate for a 100 ton/hour plant is 30 lbs/hour (approximately 0.1 gr/dscf).

Since asphalt concrete plants can burn various types of fuel and not affect product quality, they have been users of waste oils and other blended waste derived fuels. A 1985 study (Reference 34) reports that about 80% of all the purchased waste oil derived fuel in California were burned in 1984 by industries in the Standard Industrial Code 2951 - Paving Mixtures and Blocks. The majority of this code index group are asphalt concrete plants.

Compatible Wastes

Most asphalt dryers could easily be adapted to fire a liquid hazardous waste since most are oil - or combination gas and oil - fired units. The major compatibility problem would likely be the potential equipment corrosion associated with firing highly halogenated wastes. Halogen composition specifications would be required to protect the dryer and associated air pollution control equipment.

Little adverse impact on product quality is anticipated from firing hazardous waste. Organic compounds would likely be destroyed (see test results below) sufficiently by the combustion process that they would not significantly alter the product composition and quality. Metals entering the aggregate from the firing of hazardous waste will be insignificant, with a few exceptions, compared to the levels naturally occurring in the aggregate as discussed in the test results presented below.

CHARACTERIZATION OF FURNACES FOR DISPERSION MODELING

The following descriptions of furnaces, which later will be used for dispersion modeling, were assembled from data taken from a number of sources and are intended to be representative of the industry recognizing that the selected plants may not exist anywhere. They are intended to provide a basis for analysis of the environmental impacts of burning hazardous wastes on specific segments of the industry. The emission and stack data were used in atmospheric dispersion models for the risk assessment and the cost analyses. Only the furnace facilities are described; the remainder of the plant facilities are not affected by burning hazardous waste except in a very limited manner (i.e., coal mills might operate less and result in lower maintenance and operating costs). For selection of the lime kiln and the lightweight aggregate kiln, this analysis utilized the same facilities used by EPA in its NSPS analysis. The other furnaces were selected to be representative of the industry.

Wet Process Cement Kiln

The typical wet kiln has a production capacity of 850 TPD for an annual production of 266,500 tons. Primary fuel for the kiln is coal, required in the amount equivalent to 5.5×10^6 Btu/ton. Using a heating value of 12,500 Btu/pound, 7% ash, and 1.7% sulfur on an as-fired basis, the kiln will burn 7.5 tons/hour of pulverized coal. The kiln itself is 500 feet long, 11.5 feet in diameter, and rotates at 90 revolutions per hour (rph). Solid residence time is 2 to 2.5 hours.

The exhaust gases from the kiln contain 25% (by volume) water vapor, 29% CO₂, 455 ppm SO₂, 265 ppm NO_x, and 100 ppm CO. The gas discharges at 520°F into a multiclone at the rate of 150,300 acfm (60,720 dscfm). Following the multiclone is a 4-field, 2-chamber ESP which removes 98.4% of inlet dust loading and the stack discharges 137,700 acfm at 438°F (60,720 dscfm). Table 3.8 summarizes the kiln stack emission parameters used for the health effects modeling.

TABLE 3.8

EMISSION CHARACTERISTICS OF A REPRESENTATIVE
WET PROCESS CEMENT KILN

Parameters		Values
Stack Height		170 feet
Stack Diameter		11 feet
Stack Gas Flow Rate		137,700 acfm 60,720 dscfm
Stack Gas Temperature		438°F
Emission Rates		
Particulate Matter	0.11 gr/acf	130 lbs/hour
Sulfur Dioxide	455 ppm	365 lbs/hour
Nitrogen Oxides	265 ppm	93 lbs/hour
Carbon Monoxide	100 ppm	35 lbs/hour

Dry Process Cement Kiln

The configuration for the model dry process kiln is selected both to represent a stack/control device system different from the wet process plant and to approximate a typical dry process; a 4-stage suspension preheater and a rotary kiln 9 feet in diameter and 300 feet long. Capacity of the kiln is 394,000 TPY or 1,270 TPD of clinker. As with the wet process kiln, primary fuel is coal which is fed to the kiln at 6.4 tons per hour based on a heat requirement of 3×10^6 Btu/ton clinker and coal heating value of 12,500 Btu/lb, 7% ash, and 1.7% sulfur.

The kiln exhausts 154,900 acfm (104,000 dscfm) of gas at 320°F to a cyclone/fabric filter particle removal system. The control system removes 99.8% of the total dust load and discharges 15 lbs/hour of particulate matter through a 9-foot diameter stack at 120 feet above the ground. These data and other emission information are summarized in Table 3.9.

Lime Kiln

Basically, the model kiln presented here is the same configuration used by EPA in its analysis of the lime industry for the NSPS. The industry has not changed significantly in terms of size of technology since that review in 1977 (Reference 17).

The model lime kiln has a capacity to produce 500 TPD of quick lime from 1,000 TPD of limestone. Fuel is 3% sulfur coal with a heating value of 12,500 Btu/lb, and for a requirement of 7×10^6 Btu/ton of lime, is burned at a rate of 140 TPD. Dust loss from the kiln is considered to be 10% of the kiln feed rate or 8,300 lbs/hour.

The kiln exhausts this dust in 41,000 dscfm of flue gas at a temperature of 700°F with 14% (volume) water vapor. The gas contains a maximum of 1500 ppm sulfur oxides, about 150 ppm NO_x , and less than 50 ppm CO.

Particulate matter removal is provided by fabric filter which exhausts 300°F gas to the atmosphere through a short 3-foot diameter stack on top of the baghouse. Discharge elevation is 80 feet (Table 3.10).

Lightweight Aggregate

The model lightweight aggregate kiln presented here is the same as that used by EPA in its analysis of the lightweight aggregate industry for NSPS (Reference 21). This kiln has a production capacity of 1000 cubic yards per day (840 tons/day). It is fueled by pulverized coal at a rate of 157.5 tons/day for an assumed heat requirement of 4.5×10^6 Btu/ton of aggregate. The coal has a heating value of 12,000 Btu/lb and a sulfur content of 4%.

The model kiln has the following emission characteristics. Uncontrolled particulate emissions from this kiln are 842 lbs/hour. A wet scrubber is used to reduce these emissions by 99% to 8.42 lbs/hour. Uncontrolled SO_2 emissions are 238 lbs/hour. There is essentially no removal of SO_2 by the wet scrubber.

TABLE 3.9

EMISSION CHARACTERISTICS OF A REPRESENTATIVE
DRY PROCESS CEMENT KILN

Parameters		Values
Stack Height		120 feet
Stack Diameter		9 feet
Stack Gas Flow Rate		154,900 acfm
		104,000 dscfm
Stack Gas Temperature		320°F
Emission Rates		
Particulate Matter	0.011 gr/acf	15 lbs/hour
Sulfur Dioxide	520 ppm	540 lbs/hour
Nitrogen Oxide	310 ppm	140 lbs/hour
Carbon Monoxide	100 ppm	45 lbs/hour

TABLE 3.10

EMISSION CHARACTERISTICS OF A REPRESENTATIVE LIME KILN

Parameters		Values
Stack Height		80 feet
Stack Diameter		3 feet
Stack Gas Flow Rate		67,000 acfm 41,000 dscfm
Stack Gas Temperature		300°F
Emission Rates		
Particulate Matter	0.073 gr/acf	42 lbs/hour
Sulfur Dioxide	1500 ppm	615 lbs/hour
Nitrogen Oxides	150 ppm	44 lbs/hour
Carbon Monoxide	50 ppm	9 lbs/hour

The exhaust gas from the scrubber is vented to the atmosphere through a 105-foot stack that has a 4-foot inside diameter at the exit. This gas leaves the stack at 160°F and a flow rate of 49,980 scfm (Table 3.11).

Blast Furnace

The model blast furnace selected for this study is 102 feet tall with a hearth diameter of 27 feet and normally produces 3,000 tons of molten iron per day. The coke rate used is 900 pounds per ton of hot metal. This is supplemented by 150 pounds of fuel oil per ton of hot metal. Liquid hazardous waste may be substituted for the fuel oil in any proportion for purposes of the regulatory impact analysis. Hot blast air is provided at the rate of 86,000 standard cubic feet per minute and at a hot blast temperature of 2,000°F. During the production, 100,000 standard cubic feet per minute of off-gases are generated.

The off-gases from the furnace contains 25.4% carbon monoxide, 12.5% carbon dioxide, 3.5% hydrogen, on a dry volume basis. Moisture is 1.45% on a volume basis. These gases leave the blast furnace and pass through a dust removal system consisting of mechanical collectors followed by a venturi scrubber. The gases leaving the cleaning system has dust concentration of 0.01 grains per cubic feet.

Forty-one percent of the blast furnace off-gases are used to fire the blast air stove with the remainder being used to fire a boiler. Assuming that 15% excess air is used in combusting the 100,000 scfm of off-gases, 160,000 scfm of combustion products would be generated; 65,600 scfm in the blast air stoves and 94,400 scfm in the boiler. The 65,600 scfm of combustion products generated in the stoves are exhausted to the atmosphere at an average temperature of 500°F through a 200 foot stack that is 4 feet in diameter. The boiler combustion products (94,400 scfm) will exit through a 100-foot stack at 350°F. This stack is 10 feet in diameter.

Open Hearth Furnace

The model open hearth furnace selected for this study has a capacity of 320 tons of steel per heat with a tap-to-tap time of 8 hours. Thus, it can produce 960 tons of steel per day. It is fueled by natural gas and oil in approximately equal quantities on a heating value basis. The heat requirement for steel production in the furnace is 3.5×10^6 Btu per ton of steel produced.

The model furnace has the following emission characteristics. Uncontrolled emissions from the furnace is 16 pounds per ton steel produced or 640 pounds per hour. An electrostatic precipitator is used to reduce these emissions by 98% to 12.8 pounds per hour (0.02 gr/dscf).

The exhaust gas from the electrostatic precipitator is vented through a 150 foot stack that has an 8-foot inside diameter at the exit. This gas leaves the stack at 400°F and a flow rate of 65,000 standard cubic feet per minute.

TABLE 3.11

EMISSION CHARACTERISTICS OF A REPRESENTATIVE
LIGHTWEIGHT AGGREGATE KILN

Parameters		Values
Stack Height		105 feet
Stack Diameter		4 feet
Stack Gas Flow Rate		67,200 acfm 49,980 dscfm
Stack Gas Temperature		160°F
Emission Rates		
Particulate Matter	0.015 gr/acf	8.4 lbs/hour
Sulfur Dioxide	475 ppm	238 lbs/hour
Nitrogen Oxides	150 ppm	54 lbs/hour
Carbon Monoxide	50 ppm	11 lbs/hour

Spent Acid Recovery Furnace

The model spent acid recovery plant has been developed with the following characteristics:

- o Production Rate: 720 tons/day of sulfuric acid
- o Fuels: 252 tons/day sulfur total
84 tons/day elemental sulfur
126 tons/day spent acid
16,800 tons/day HWDF with 50% S and 50% organics
- o Exhaust Temperature: 90°F
- o Exhaust Flow: 20,000 scfm
- o Stack Height: 75 feet
- o Stack Diameter: 4 feet
- o Air Pollution Controls: packed tower
dual absorption system
- o Emission Rate: .08 gr/dscf particulates at 7% oxygen
4 lbs/hr HCl
99.99% DRE

Asphalt Aggregate Kiln

The model asphalt concrete plant selected for the engineering cost analysis and the risk assessment is as follows:

- o Plant Type: batch-mix plant
- o Production Rate: 200 tons per hour
- o Fuel: 300 gallons per hour of fuel oil or waste derived fuel
- o Operational Rate: 5 hours per day
5 days per week
40 weeks per year
- o Stack Characteristics
 - Temperature: 150°F
 - Flow rate: 20,000 scfm
 - Moisture content: 25%
 - Diameter: 4 feet
 - Height: 25 feet
- o Air Pollution Control: low energy scrubber (spray bars, damper, wet fan)

o Emission Rates:

.08 gr/dscf particulates at
7% oxygen
4 to 30 lbs/hr HCl
99.99% DRE

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SECTION 4

DESTRUCTION AND REMOVAL EFFICIENCY OF HAZARDOUS MATERIALS BY INDUSTRIAL FURNACES

The EPA Office of Solid Waste (OSW) recognized the need to collect emission data from industrial furnaces burning hazardous wastes as fuels to support its regulatory efforts. EPA therefore developed a program to test industrial furnaces burning a variety of waste streams. This section describes the analytical methods used to test stack gases from furnaces burning hazardous wastes, identifies those testing procedures that are or should be standardized for collecting test burn data, evaluates the QA/QC activities that are or should be required for each method and identifies the quality acceptance criteria that were applied to test burn data collected from seven kilns, three asphalt aggregate kilns, and a blast furnace.

Results from eleven trial burns (ten EPA and one California Air Resources Board test) were analyzed to determine if the destruction of hazardous materials was adequate to protect the environment. Included in the analyses and discussion are: the DRE, the POHCs, emissions of particles, emissions of HCl, PICs, metals emissions, and combustion gases.

The last portion of Section 4 describes the types of system modifications that would be required of existing production furnaces to enable them to safely burn hazardous wastes as fuels. Major facility modifications could include:

- o Installing waste storage and handling facilities
- o Installing laboratory testing facilities
- o Installing waste pretreatment facilities
- o Modifying or adding burner equipment
- o Modifying or adding combustion controls
- o Installing monitors for waste feed rate and stack gases (CO and O₂)
- o Modifying or adding air pollution control equipment

EVALUATION OF TEST METHODS FOR MEASURING ORGANIC EMISSIONS FROM COMBUSTION SOURCES

The most widely employed procedures for measurement of the rates of emissions of organic substances from combustion sources are the Modified Method 5 (MM5), the volatile organic sampling train (VOST) procedures, and EPA Method 23. The first two methods are conceptually similar; both sampling trains consist of a particle filter, a condenser, a bed of porous polymer sorbent, and a condensate trap. Their differences lie in their size, sorbent, and analytical techniques. Some of the salient attributes of the two methods are compared

in Table 4.1. Method 23 specifies that stack gas samples be drawn into inert plastic bags at a constant rate by a lung-type sampler. Analyses are usually done in the field by GC with either flame ionization or electron capture detection. The strengths and weaknesses of these test methods are discussed in this section.

Modified Method 5 (MM5)

This method is an adaptation of EPA Method 5 (40 CFR Part 60) modified to obtain samples for organic compound analysis as well as quantification of particulate matter emissions. The adaptation (Figure 4.1) is the addition of a sample gas condenser and a sorbent resin module between the heated filter and first impinger of Method 5 train. The sorbent resin most commonly used for hazardous waste combustion evaluation is XAD-2 which is highly effective at trapping organic compounds with boiling points greater than 100°C (sometimes referred to as semi-volatile organic compounds or S-VOC).

The sample is collected by isokinetically drawing stack gas through a heated glass or quartz probe, through a heated glass fiber particle filter and then to the condenser/resin module. The sample gas is kept above 120°C until it reaches the condenser where it is cooled to <20°C. Filtration temperatures up to 150°C are used to minimize organic species condensation prior to the condenser if this does not interfere with the determination of particulate matter. The sample gas and condensate pass through a resin bed located below the condenser allowing the condensate to percolate through the bed and collect in an impinger or condensate trap underneath the resin module. The sample gas is then bubbled through two more impingers in the conventional Method 5 configuration for acid gas and additional moisture removal.

Samples are analyzed by performing solvent extractions on the probe and filter material as one fraction, the resin as a second fraction, and optionally on the condensate and impinger catches as third and fourth fractions. The solvent extracts are concentrated and may be combined and analyzed by gas chromatography (GC) using mass spectrometers (MS), flame ionization detectors (FID), or electron capture detectors (ECD), as appropriate, for the organic compounds of interest.

MM5 Procedure Standardization

The construction and operation of a Method 5 train is well known and well described in the literature. The train operation, sample recovery, choice of sorbent resin, and analytical method to be used in a given application of MM5 are not specified. At this time, a single description of how the MM5 train is or may be used is not available. The sample collection aspects including resin choice, are discussed in various publications, but the sample recovery, analysis, and data reduction are not well described.

The Method 5 train and sample collection scheme has become the standard for measurement of emissions of particulate matter and it is not surprising that modifications to it have evolved in attempts to quantify other types of emissions using the same equipment and techniques. Method 5 is written as a test procedure for determining compliance with New Source Performance Standards (NSPS). This specificity has not been established for the MM5 procedure as applied to hazardous waste DRE sampling in part because at least some

TABLE 4.1
COMPARISON OF MM5 AND VOST PROCEDURES

Feature	MM5	VOST
Sampling Rate	14-40 lpm	0.5 to 1 lpm
Sorbent	XAD-2	Tenax®-GC
Analysis of Condensate	Yes	No
Sample Recovery Technique	Solvent Extraction	Thermal Desorption
Analysis	GC or GC/MS	GC/MS
Boiling Point of Analytes	>100°C	30° to 100°C
Sampling Duration	1 to 4 hours	0.3 to 1 hours

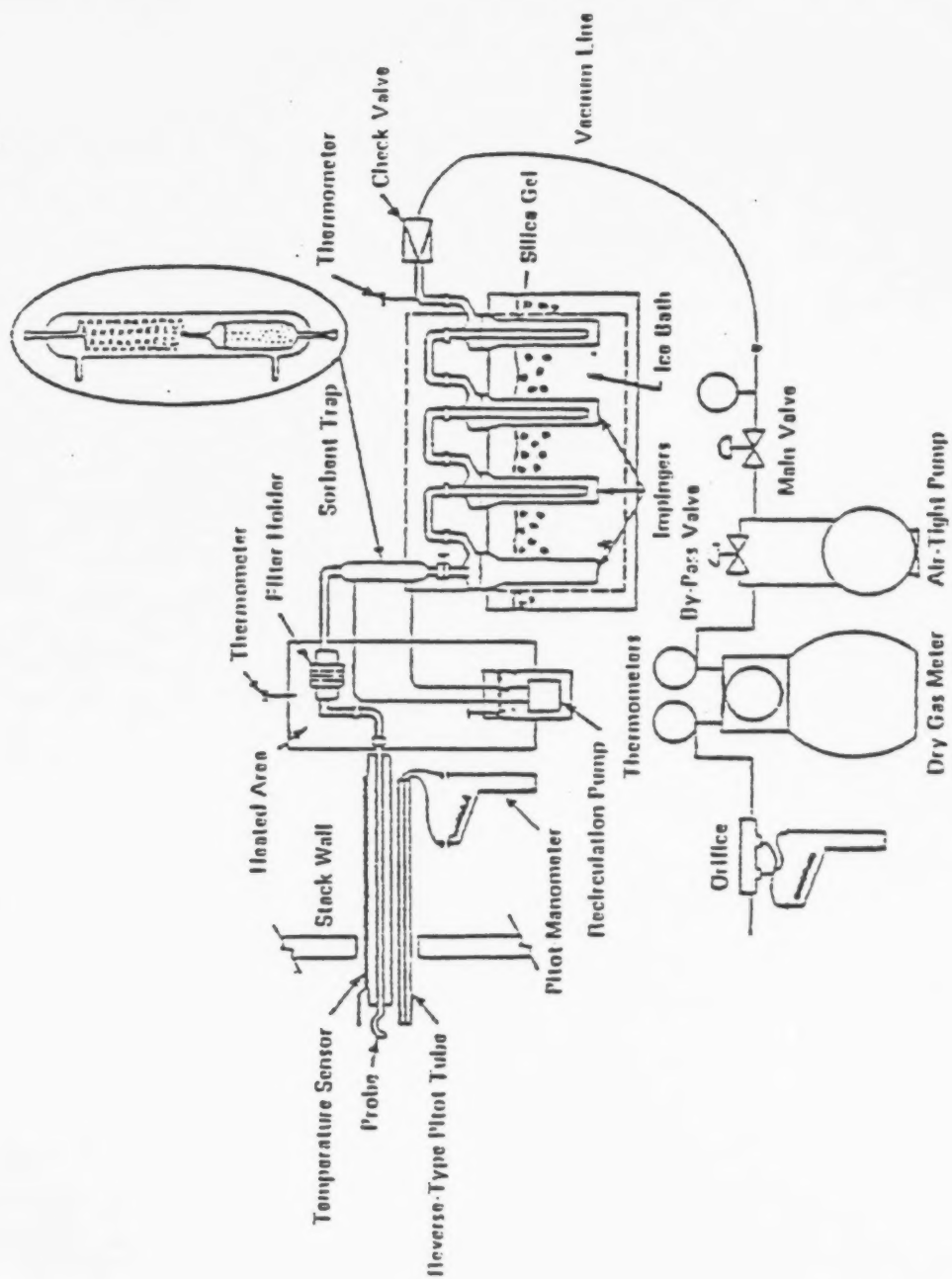


FIGURE 4.1
 MODIFIED METHOD 5 TRAIN (MM5)

of the measurement objectives will be different from test to test. Many of the objectives will be the same and these need to be identified and the method written to ensure that the goals are clear.

First, the relative priorities of measuring particulate matter and organic compound emissions need to be set. Most often these will not conflict and no sacrifice of one measurement for another is needed. In the cases where some relatively significant quantity of organic chemical of interest may be found in the "front half" (probe, filter, connecting glassware) of the Method 5 train then the accuracy of particulate matter measurement may need to be sacrificed to obtain the S-VOC. Examples are not brushing the probe to avoid contamination, removing some particulate matter with the solvent used for S-VOC recovery, and/or operating the heated portion of the train at higher than normal temperature (e.g., 205°C) and vaporizing particulate matter which would otherwise deposit in the front half of the train. In the case of oil-fired boilers, much of the particulate matter is carbonaceous with little inorganic ash and the material contains straight chain and aromatic hydrocarbons which may or may not be collected in the front half depending on the filtration temperature. This is different from a coal-fired unit where the ash is mostly (typically >95%) inorganic matter. If POHCs or PICs to be measured from hazardous waste combustion are to include naphthenic or paraffinic hydrocarbons, then whether the boiler is oil- or coal-fired will probably affect the sampling and analysis scheme.

What compounds are selected as POHCs and PICs also bears on the choice of a suitable sorbent resin. Generally, the MM5 is used to sample for PICs - the higher boiling compounds, but it is also used to collect samples for certain Appendix VIII compounds, e.g., toluene, monochlorobenzene. XAD-2 is the resin most commonly used and recommended as a general purpose sorbent when solvent extraction is the means of sample recovery. XAD-2 was selected for its sorbent properties, ease of cleaning, and sample recovery efficiency. Others are also used, for example, Tenax®-GC, if thermal desorption is the sample recovery procedure. However, selection of a sorbent resin usually involves a great deal of time and effort (literature and laboratory research) that is usually not practical on a case-by-case basis.

The last major uncertainty in regard to applying the MM5 DRE testing is analysis - compound identification and quantification. Currently, GC is used with one of three detection modes MS, ECD, and FID. Each has its own benefits and disadvantages relating to sensitivity, reproducibility, compound identification, interference rejection, and analytical cost. Each of these factors needs to be considered and the method written to describe when each would be appropriate.

MM5 QA/QC Evaluation

Most of the QA emphasis has been on obtaining acceptable blank values and preventing contamination. The large gas volume and relatively large quantity of resin concentrates the sample, which in conjunction with the solvent extraction, usually results in a greater analyzable mass than methods employing a lower sample volume. This greater analytical mass tends to decrease the importance of trace contamination of the sample. The resin cleaning, blank extraction, field trip, and laboratory handling blanks are adequate to identify problems and likely causes.

There is one important aspect lacking in the method as it is being used and this is use of field spikes to check on sample loss and recovery efficiency. As currently practiced, the procedure does not provide a means for determining target compound collection efficiency or for evaluating sample recovery. The nature of the sample and the field conditions preclude the usual option of splitting the collected sample and spiking one split with a target compound to check loss and recovery. These spikes and replicate analyses can be performed in the lab using the extractate with some loss of sensitivity. It is also possible, and it seems highly advisable, to spike the field samples with a tracer compound having properties similar to the target compounds.

Because the volume of gas sampled by MM5 is large, the total quantities of the various S-VOCs in the samples ranged from a few to several hundred micrograms. Thus, the MM5 results are not greatly influenced by even a few hundred nanograms of contaminants.

Two QA acceptance criteria were applied to the MM5 data before they were included in this document.

1. The recovery of surrogate or spike compounds added to the sample before analysis must have been in the range of 50 to 150%.
2. The rate of feed of any given compound must have been 10,000 times the minimum detectable limit of the MM5 procedure.

The first of these is merely a demonstration of acceptable analytical accuracy. The acceptable range is wide relative to normally attainable analytical precision. It is adequate for this analysis because the DRE calculation is insensitive to an error of a factor of three in the emission rate measurement.

The calculation of DRE, the primary use of the data in this document, is based upon the concentration of the various constituents in the waste feed stream and in the exhaust gas. The accuracy and precision of these concentration measurements decrease when they are near the limit of detection of the analysis methods. Consequently, the dispersion of DRE values that are calculated based upon these imprecise measurements becomes unacceptably large. The decision was made to include only those data that are as accurate as is possible using the available methods. The method chosen, the second of the QA acceptance criteria, accomplishes this goal.

Volatile Organic Sampling Train (VOST)

The basic details of construction and operation of the VOST are described in the "Protocol for Collection and Analysis of Volatile POHCs Using a Volatile Organic Sampling Train (VOST)" by Envirodyne Engineers for IERL (Reference 1). The highlights of the procedure are described below. Stack gas is drawn through a quartz wool particle filter in a glass or quartz probe heated to approximately 130°C, through a three-way stopcock and through a coil condenser. Following the condenser, the sample gas passes through a glass tube containing 1.6 g of Tenax®-GC, a condensate trap, a second condenser, and a second sorbent tube containing 1.0 g Tenax®-GC followed by 1.0 g of activated charcoal. A second condensate trap is next, followed by valves, flow meters,

gas meters, etc. All portions of the sample line preceding the last condenser are glass, stainless steel, or Teflon®. Figure 4.2 is a schematic depiction of the train. Previous experiments have established 20 liters as a maximum safe sample volume for this train. A greater sample volume incurs the risk of stripping sorbed POHCs off the resin. The sample rate for these tests was 1/2 liters per minute for a total of 40 minutes per pair of tubes.

An ice water bath is used to circulate water through both condensers to maintain sample gas temperature below 20°C through the sorbent tubes. Temperature of the probe liner, first condenser outlet, ambient air, and dry gas meter are measured and recorded. Leak checks of the whole train and each pair of sorbent tubes for each run are conducted and the resulting vacuum is released by allowing ambient air in through a charcoal filter connected to the three-way valve.

Tubes used for the kiln tests were of the inside-inside design that are held in the sample train with stainless steel Swagelok® fittings and ceramic-filled Teflon® ferrules. Other samplers have used the inside-out-inside design; a double walled sorbent cartridge/shipping container that uses O-rings and end caps to seal the cartridges. Stainless steel caps are used to seal tube ends for shipment before and after sample collection. After sample collection, tubes are kept and shipped in chilled styrofoam containers.

The tubes were analyzed on a GC/MS using thermal desorption with trap and purge. The method is described in the protocol and involves spiking each tube or pair of tubes with an internal standard, thermally desorbing the tubes into a water trap, and purging the water trap onto an analytical column for component separation. Identification and quantification are made by elution times, characteristic ions, and ion current profile using a computerized data library.

VOST Standardization of Procedure

The protocol provides clear and specific directions about the sample train to be used and the method of sampling and the method of analysis. Reagent preparation, sample handling, QA/QC activities, calibration and calculations are all described in detail. The protocol states that conditioned cartridges, as well as used ones with sorbed sample, be kept in ice water before use and after sample collection. This ice water storage is not required if acceptable blank levels can be maintained.

Options regarding sample collection and recovery (analysis) efficiencies are also provided along with evaluation criteria. The analytical procedure is also described very specifically.

VOST QA/QC Evaluation

The pre-sampling QA activities are clear and direct. Tenax® and charcoal cleaning, tube packing, and desorption blanking, provide sufficient assurance that the sample cartridges start clean. The trip blank, field blank, and lab blanks are intended to provide a history and background levels of contamination and/or degradation so that the results of the sample analysis reflect only POHCs present in the stack gas. This history is especially difficult to create if the samples and blanks are not analyzed promptly. Sample

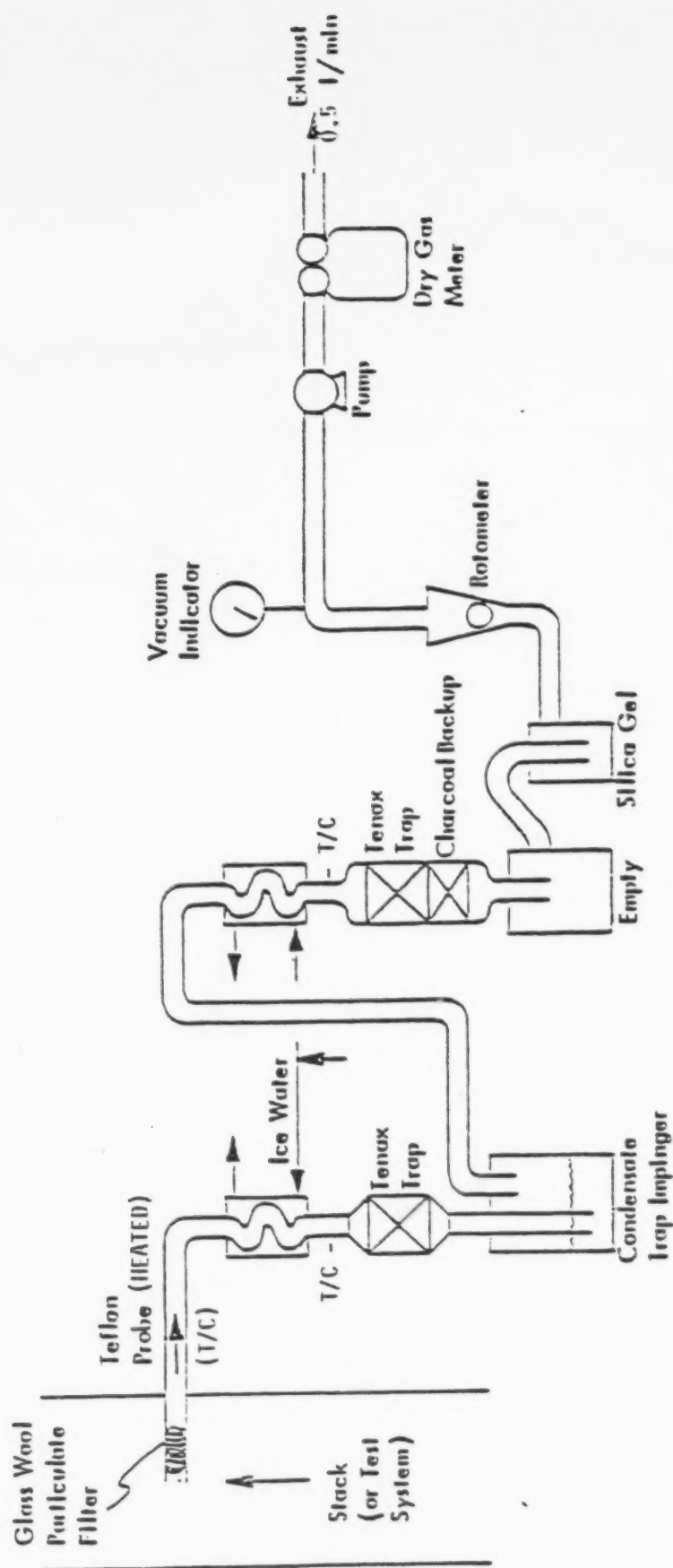


FIGURE 4.2

Volatile organic sampling train (VOST).

degradation, tube cross-contamination, contamination from external sources (lab air, ambient air, etc.), and calibration and response standard degradation become more likely and less distinguishable.

As each analysis is a one time occurrence (no way to split), a bad analysis, a contaminated or otherwise invalid tube, means a lost data point. However, the sample collection involves three pairs, so duplication is inherent in the sampling. One could carry this suggestion further to say that two blanks would be desirable to yield an average value and increase the confidence in the results.

The protocol requires one exposure pair per six pairs of sample tubes. These exposure (field or shadow) blanks are opened as if they were sample tubes but are not installed in the train.

One QA action which has not been done, for several reasons, is spiking tubes in the field with one or more target POHCs or surrogates to establish recovery efficiencies. The difficulty stems from two major areas. One is the difficulty in maintaining reagent and tube purity in a field environment and the other is not being able to split a single sampling and spike one portion. The methods used to date have centered on lab simulations and the analytical process. A suitable field spike procedure would yield data on sample loss (leakage out) and desorption efficiency as well as additional data on contamination, lab QA, and overall method validity.

The following list identifies areas where additional effort and investigation could strengthen the VOST procedures.

- o Spike blank cartridges in the field with a labeled compound to detect potential leakage during field storage and subsequent transport.
- o Analyze the sample immediately with as little storage time in field and lab as possible.
- o Conduct a detailed systematic evaluation of field handling, field storage, shipping, and lab storage, to identify potential causes of contamination and/or leakage. Develop guidelines to detect and prevent contamination and to leak check cartridges.
- o Investigate the current seal design. Do temperature changes cause leakage? How can one be certain the tubes are perfectly sealed? How can overtightening/undertightening be prevented? Can a pressure/vacuum tight seal be obtained repeatedly on a large number of tubes with no failures?
- o Investigate cross-contamination. Place spiked samples and clean blanks in the same container, store one or two weeks, and analyze. Do the above with loose fittings or cracked tubes and observe for cross contamination. Place the samples and lab blanks with their double seal in an atmosphere containing trace amounts of methylene chloride or waste fuel vapors, store, and analyze.

QA acceptance criteria were developed for the VOC data produced during each test. These criteria could not be identical for all sites since the methods used differed. In all cases, only those components of waste listed in Appendix VIII (CFR 40 Part 261) were included in the DRE results; even though there may have been other organic constituents measured in the waste feed and stack exhaust streams.

It is noted that not all of the VOC sampling done during these tests was done by the VOST procedure as has been described. The train used to sample VOC at Site A consisted of a condensate trap and one Tenax®-GC trap in series. No condenser or sorbent temperature control was used at Site A. Impingers were inserted in the train upstream and downstream of the Tenax®-GC tube at Site C.

The VOC sampling trains used at Sites A and C varied from the VOST train in that an impinger (containing water) in an ice bath was inserted in the sample line ahead of the Tenax®-GC cartridge. Three QA acceptance criteria listed below were applied to the data collected by these trains:

1. If impingers were used, the contents must have been analyzed.
2. Both sorbent tubes must have been analyzed.
3. At least 70% of the total quantity of any compound found on the sorbent tubes must have been found on the first (Tenax®-GC) tube.

The first two of these are completeness criteria. While analysis of the condensate is not normally a part of the VOST protocol, it is necessary in these instances because of the location and the temperature of the condensate trap. The volume of condensate obtained from a 20-liter sample of stack gas is approximately 1.5 ml. Even though the compounds of interest (mostly chlorinated hydrocarbons) are normally considered to be insoluble in water they are miscible to a small but measurable extent. A compound soluble to 1 mg/liter is said to be insoluble yet that translates to 1500 ng/1.5 ml of condensate which is large relative to the analytical quantities of interest. The VOST train causes the condensate to be drawn through the resin bed. The resin should remove the compounds from the condensate. Since there was no contact between the condensate and the resin at these sites, it was necessary to analyze the condensate.

The third acceptance criteria was included to eliminate contaminated samples from the data. Persons who have sampled surrogate stack gases spiked with chlorinated hydrocarbons under controlled conditions have reported that at least 90% of these compounds are sorbed on the first resin trap. This is not true of highly volatile compounds, e.g., vinyl chloride, but it is for the compounds of interest in this document. Tenax®-GC has sufficient affinity for these compounds to remove them nearly quantitatively from the sample gas stream. The charcoal, used as a back-up sorbent in the second cartridge, has a much greater sorbent capacity and affinity for these compounds. Thus, if the cartridges are exposed to contaminants, the second tube should sorb them at a higher rate. Therefore, setting the acceptance criteria at 70% allows acceptance of some contamination but rejects grossly contaminated cartridges.

At the remaining sites (B, D, and F) the VOST train (Figure 4.2) was used. Acceptance criteria #1 is not applicable but the other two are. Note

that not all of the test reports contain sufficient data to allow comparison of the results to the QA acceptance criteria. These data have been accepted with the expressed reservation that their quality is unknown. The effects of application of these QA criteria on the data are discussed for each test in the following pages.

EPA Method 23 (M23)

M23 was proposed by EPA on June 11, 1980, but has not been promulgated as a reference method. It is a method for determination of the concentrations of low boiling halogenated organic compounds. It is not applicable to compounds that might condense at ambient temperatures, are sorbed on particles or are water soluble. It is also not applicable to measurement of oxygenated compounds (alcohols, ketones) as these appear to be rapidly adsorbed by the sample container.

A sample of gas is collected in an inert (Tedlar® or aluminized Mylar®, flexible walled bag. The sample is conveyed directly into the bag from the stack through Teflon® tubing. The bag is placed inside a rigid walled, leak tight container, evacuation of this container causes expansion of the bag which draws sample gas into the bag. This system eliminates the need for contact of the sample with the pump or other potential sources of contamination. Figure 4.3 is a schematic representation of the sampling equipment the method specifies that samples be analyzed by a GC with a FID. Some users of the method have substituted ECDs or Hall electrolytic conductivity detectors (HECD) for the FID. Both are more sensitive to halogenated compounds than is the FID. The procedure requires that the analysis be done within two hours of sample collection. In most cases, this requires that the analysis be done in the field.

M23 Standardization of Procedure

The method as it appeared in the Federal Register is clear and specific. The sampling and analysis equipment are described in detail. Clear and specific instructions are given about sampling and analysis procedures, QA/QC activities, calibrations, and calculations. The method should be updated to include the HECD and the ECD in order to increase its sensitivity and specificity.

QA/QC Evaluation

The procedure describes two alternate calibration procedures: one employs cylinder gases with known concentrations of chlorinated hydrocarbons, the other employs flash evaporation of known quantities of compounds. Both require use of the bags during the calibration procedure which provides some assurance that the sample bags to be used are not unusually sorptive of the compounds of interest. An additional QA step, spiking stack gas samples, would provide assurance that water and other stack gas constituents do not interfere.

The calibration section of the method describes the possibility of cross-contamination if a low concentration standard is prepared in a bag that has contained a high concentration standard. The same cross-contamination between stack samples and between standards and stack samples is possible. There is

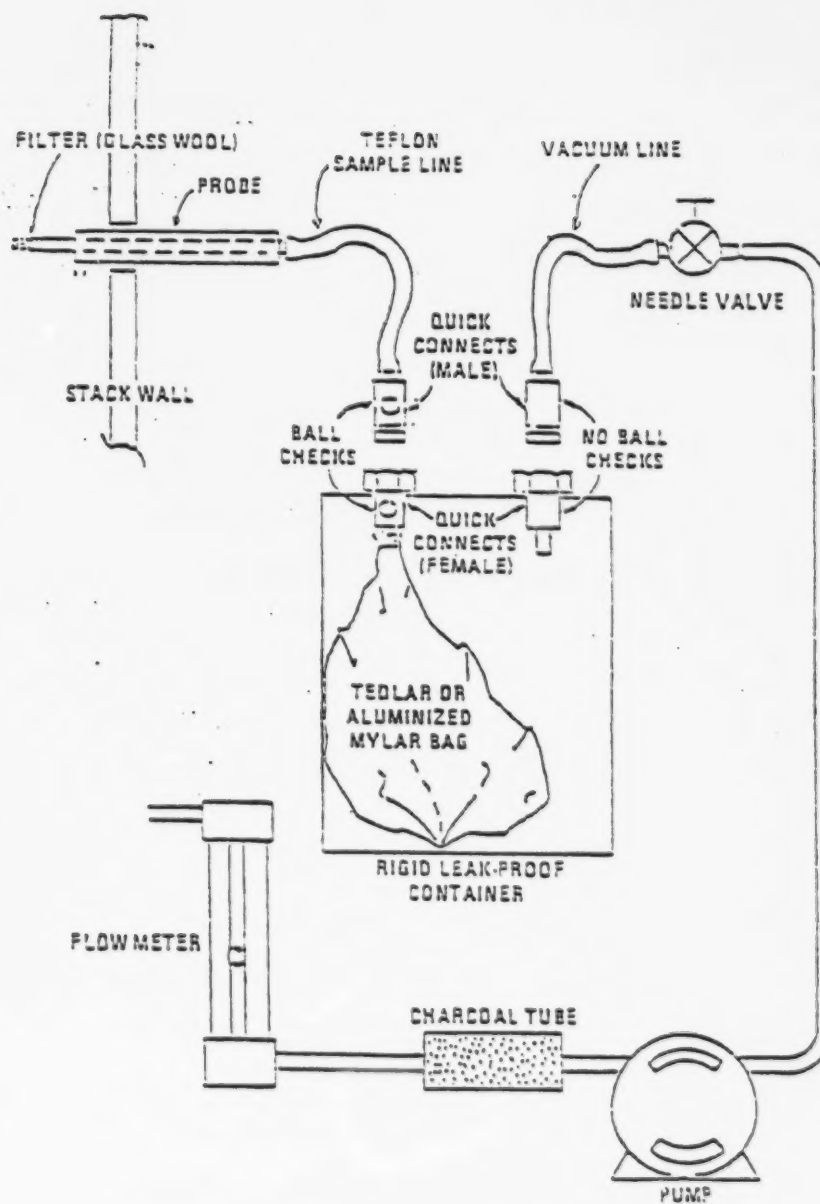


FIGURE 4.3

METHOD 23 SAMPLING TRAIN

also the possibility that the bags may have been contaminated during manufacture or during previous use. These possibilities should be addressed in the QA procedures. This step could be incorporated easily by requiring that the leak check be done with pure gases and that the gas be checked for contaminants prior to use of the bag. A similar step could be mandated between separate uses of a bag.

Verification of the cleanliness of each bag prior to its use is required so trip, field, and laboratory blanks are not appropriate. The requirement for analysis of the samples within two hours of their collection makes sample splitting impractical. The method could be made more useful by validating the stability of more compounds in the inert sampling bags. The method is applicable to more than the seven compounds that the method lists but the tester must perform a test specific validation study if it is to be used for compounds not listed in the method.

Results of the test burns conducted to support the regulatory program for industrial furnaces are evaluated in the following paragraphs. The evaluation is presented in three separate categories by type of industrial furnace. Cement, lime, and lightweight aggregate kilns are discussed as a single category because these furnaces share many common performance characteristics. The three tests for asphalt aggregate kilns are presented as a separate category as is the single test for blast furnaces.

EVALUATION OF TEST BURN RESULTS FOR CEMENT, LIME, AND LWA KILNS

The initial surveys that were performed during the early stages of development of this BID revealed that adequate information about the DRE of hazardous compounds by kilns did not exist. EPA has also developed extensive data that reported the DRE (DRE includes both thermal destruction and removal by control devices) of hazardous compounds by incinerators. There are significant differences in the two processes. Incinerators typically hold their combustion gases in an oxidizing atmosphere at temperatures ranging from 1800° to 2500°F for times ranging from 2.0 to 3.0 seconds. The combustion zone temperature in kilns is typically higher (2250° to 3000°F) and the retention time is typically as long or longer (2 to 5 seconds). Kinetic theory predicts that elementary reactions should be faster at the higher kiln temperature, faster by a factor ranging from 4 up to 20,000, depending upon the activation energy of the particular reaction. This range has been confirmed by thermal destruction analytical system (TDAS) data for many common hazardous compounds. These data demonstrate that rates of destruction increase by factors that range from 17 to 12,000 when the temperature is raised from 1900° to 3100°F (References 3, 4, and 5).

EPA undertook a series of tests to determine whether the destruction of hazardous materials by co-firing in kilns was adequate to protect the environment. Beginning in October 1981, EPA performed tests at six kilns. An additional test, performed by the State of California Air Resources Board, has been included in this data set. The sources that have been tested are characterized in Table 4.2. Three dry process cement kilns, two wet process cement kilns, one lime kiln, and one lightweight aggregate kiln are included.

TABLE 4.2

COMPLETED FIELD TESTS ON HAZARDOUS WASTE CO-FIRED IN KILNS

Site Designation	Kiln Type/Size	Primary Fuel	Waste Description	Source Emission Control	Sampling Matrix
A	Dry Process Cement (150,000 lb/hr or 1,800 TPD)	Pulverized Coal (low sulfur)	Mixed solvents	Baghouse	- o Flue gas -- Modified Method 23 for volatile organic compounds -- Tenax®-GC train for volatile organic compounds -- SASS for metals and semi-volatile organic compounds and PCBs -- Method 5 for particles, HCl -- Continuous monitors for SO₂, NO_x, O₂, CO, CO₂, and TUHC - o Waste dust - o Waste fuel
B	Wet Process Cement (60,000 lb/hr or 720 TPD)	Pulverized Coal	Mixed solvents (85% aromatic, 1% chlorinated)	Electrostatic Precipitator	- o Flue gas -- VOST for volatile organic compounds -- Modified Method 5 for semi-volatile organic compounds, particle metals -- Impingers for HCl -- Continuous monitors for CO₂, NO_x, CO, O₂, and TUHC - o Waste fuel - o ESP waste dust - o Clinker - o Kiln feed - o Coal - o Slurry water

TABLE 4.2--Continued

Site Designation	Kiln Type/Size	Primary Fuel	Waste Description	Source Emission Control	Sampling Matrix
C	Lightweight Aggregate (20,000 lb/hr or 240 TPD)	Pulverized Coal	Solvents, alcohols, ethers, still bottoms	Cross-flow Scrubber	- o Flue gas -- VOST for volatile organic compounds -- Method 5 for particles and metals -- Impingers for HCl -- EPA Method 3 for CO₂ and O₂ -- EPA Method 7 for NO_x -- EPA Method 6 for SO₂ - o Waste fuel - o Scrubber blowdown - o Product - o Raw material - o Coal - o Scrubber water
D	Lime (17,000 lb/hr or 204 TPD)	Petroleum Coke (90%) Gas (10%)	Solvents, lacquer, thinners, alcohols, still bottoms, paint wastes	Baghouse	- o Flue gas -- VOST for volatile organic compounds -- Method 5 for particles, metals, HCl -- Method 3 for O₂ and CO₂ -- Continuous monitors for NO_x, CO, SO₂, TUHC, and O₂ - o Waste fuel - o Baghouse dust - o Product - o Raw material - o Coal

TABLE 4.2--Continued

Site Designation	Kiln Type/Size	Primary Fuel	Waste Description	Source Emission Control	Sampling Matrix
E	Dry Process Cement (120,000 lb/hr or 1,440 TPD)	Coal/ Petroleum Coke Mixture	Solvent reclamation still bottoms (2-5% Cl)	Electro- static Precipitator	- o Flue gas -- Method 23 for volatile organic compounds -- Method 5 for particles and metals -- EPA Method 8 for SO₂ -- EPA Method 7 for NO_x -- EPA Method 25 for TGNMO -- Impingers for HCl -- Continuous monitors for CO, CO₂, and O₂ - o Waste fuel - o ESP dust - o Product - o Coal
F	Dry Process Cement (120,000 lb/hr or 1,440 TPD)	Coal/Coke Mixture	Hydrocarbon solvents (2% Cl)	Electro- static Precipitator	- o Flue gas -- VOST for volatile organic compounds -- Modified Method 5 for particles, metals, semi-volatile organic compounds -- Impingers for HCl -- Continuous monitors for CO₂, NO_x, SO₂, CO, O₂ and TUHC - o Waste fuel - o ESP dust - o Product - o Raw material - o Coal/coke - o Quench water

TABLE 4.2--Continued

Site Designation	Kiln Type/Size	Primary Fuel	Waste Description	Source Emission Control	Sampling Matrix
G	Wet Process Cement (70,000 lb/hr or 840 TPD)	No. 6 Oil	Degreaser and pharmaceutical wastes (6 to 35% Cl)	Baghouse	- o Flue gas -- Method 23 for volatile organic compounds -- Method 5 for particles and metals -- EPA Method 6 for SO₂ -- EPA Method 7 for NO_x -- EPA Method 9 for opacity -- EPA Method 10 for CO -- EPA Method 3 for CO₂ and O₂ -- Impingers for HCl -- SASS for semi-volatile organic compounds - o Process water - o Fuel oil - o Waste fuel - o ESP dust - o Product

The percent fuel replacement chlorine and heat value of the wastes burned at the various sites are summarized in Table 4.3. More detailed information about the waste fuels is given in the individual test descriptions.

DREs of POHCs

The data from all seven test sites (References 6 through 12) were reviewed and subjected to the QA acceptance criteria that were presented previously. The results at each site were compared to the promulgated incinerator regulation individually.

The DRE of POHCs in kilns are summarized, by site, in Table 4.4. The compounds listed are those that were measured in the waste and also listed (except Freon-113) in Appendix VIII (Reference 13). Particle, HCl, and PIC emissions are discussed in separate sections that follow the individual test discussions. Likewise, an overview of the implications of all the data is presented following the individual test discussions.

Site A

Site A was a dry process cement plant with a production capacity of 150,000 pounds of cement clinker per hour. The primary fuel used at the site was low-sulfur pulverized coal. A portion of the clinker cooler exhaust gas is used as combustion air in the kiln. The kiln exhaust gas temperature was reduced from approximately 1000°F to 575°F by a water quench. Particulate matter emissions were controlled by a 14-compartment baghouse. Normally, approximately 100 tons per month of the dust collected by the baghouse are removed from the system for disposal. This waste rate is sufficient to maintain an acceptably low alkali content in the clinker product. During these hazardous waste combustion tests, the disposal rate was increased to 2000 tons per month. The increased rate of disposal was necessary to remove the excess chloride that the waste introduced into the system. The waste, mixed industrial solvents, was supplied by a waste fuel broker. The waste was approximately 30% alcohols and ketones, 40% aromatic compounds, and 20% aliphatic compounds. It contained approximately 2% chlorine. Waste fuel provided approximately 30% of the heat input during the co-fired runs.

A total of 31 test runs were performed. Eighteen of these were baseline runs with the kiln firing coal only. Nine of these runs were done five months before the 13 co-fired runs and nine during the same week as the co-fired runs. Only two compounds listed in Appendix VIII were in the waste at concentrations sufficient to allow computation of DREs. They were methylene chloride (MeCl_2) and methyl chloroform (1,1,1-trichloroethane).

Two different methods were used to measure the stack gas emission rates of the two POHCs. One of these was M23, the other was a single Tenax® tube that was preceded by a condensate trap. The M23 sampling and analysis was done in accordance with the provisional method as published in the Federal Register. QA/QC data showed that the results were generally credible though a few results were discarded because of suspected contamination. The results of the Tenax® analyses also exhibited symptoms of contamination. For example, the baseline test emission rates of both POHCs were higher (Table 4.4) than the co-fired emission rates. On the average the Tenax® results were higher

TABLE 4.3

WASTE FUEL CHARACTERISTICS OF
HAZARDOUS WASTE CO-FIRED IN KILNS

Site	Primary Fuel	Fraction of Heat by Waste (%)	Percent Chlorine in Waste	Heat Value of Waste (Btu/lb)
A	Pulverized Coal	30	2	13,800
B	Pulverized Coal	40-60	1-4	12,400
C	Pulverized Coal	55	0.9	11,800
D	Petroleum Coke & Gas	8-36	3.0	12,900
E	Pulverized Coal & Petroleum Coke	13	4.5	12,300
F	Pulverized Coal & Petroleum Coke	35	1.7	12,300
G	No. 6 Oil	10	25	10,200

TABLE 4.4

DESTRUCTION OF PRINCIPAL ORGANIC HAZARDOUS
CONSTITUENTS IN KILNS

Site	Compound	Baseline Emissions (No Waste Feed) (lb/hr)	Co-Fired Emissions (lb/hr)	Average Co-Fired POHC Feed Rate (lb/hr)	Test Average DRE (%)
A	methylene chloride	0.0008	0.0003	15	99.998
	1,1,1-trichloroethane	0.0012	0.0006	30	99.998
B	1,1,1-trichloroethane	0.0004	0.0010	20	99.995
	methyl ethyl ketone	0.0070	0.0050	50	99.990
	toluene	0.0380	0.0380	100	99.962
C	methyl ethyl ketone	a	0.0008	40	99.998
	methyl isobutyl ketone	a	0.0004	20	99.998
	perchloroethylene	a	0.00004	1	99.997
	toluene	a	0.00146	146	99.999
D	methylene chloride	0.00001	0.00002	1	99.9974
	methyl ethyl ketone	0.00006	0.00008	16	99.9995
	1,1,1-trichloroethane	0.00004	0.00006	2	99.9969
	trichloroethylene	0.00002	0.00003	15	99.9998
	perchloroethylene	0.00002	0.00003	15	99.9998
	toluene	0.00013	0.00026	85	99.9997
E	methylene chloride	b	<0.0882	95	>99.909
	methyl ethyl ketone	b	<0.0738	230	>99.968
	1,1,1-trichloroethane	b	<0.135	50	>99.731
	toluene	b	<0.0936	295	>99.968
F	Freon-113	0.00308	0.00040	40	99.999
	methylene chloride	0.03389	0.00621	16	99.96
	methyl ethyl ketone	0.00133	0.00109	55	99.998
	1,1,1-trichloroethane	<0.00022	<0.00011	11	>99.999
	toluene	0.01405	0.00745	150	99.995
G	methylene chloride	0.0108	0.754 ^c	223	99.662 ^c
	chloroform	0.0886	0.836 ^c	277	98.698 ^c
	carbon tetrachloride	0.5600	2.140 ^c	126	98.300 ^c

^a No baseline run at this site.

^b No POHC detected during baseline runs.

^c Baseline emission rate has been subtracted.

than the M23 results by a factor of four. All emission rates were low, however, so that even the DREs calculated from the Tenax® method, without making allowances for the relatively high baseline measurements, were in excess of 99.99%. The Tenax® results which are given in Table 4.4, demonstrate that this kiln destroyed more than 99.99% of the POHCs in the waste stream.

Site B

Site B was a wet process cement kiln with a production capacity of 60,000 pounds of clinker per hour. The primary fuel used at this site was pulverized coal. A portion of the clinker cooler exhaust gas is used as combustion air for the kiln, the remainder either passes through cyclones prior to its use as coal suspension air or is exhausted. The dust collected by the cyclones is added to the cooled clinker prior to its grinding. Kiln emissions are controlled by an ESP. The dust collected in the first field ESP hopper (3-4 tons/hour) is recycled into the burner end of the kiln. The dust collected in the second, third, and fourth field hoppers (3-4 tons/hour) is generally high in chloride and is wasted to control the chloride content of the product.

The waste burned at this site was supplied by a waste broker. It consisted largely (27%) of aromatic solvents and contained approximately 10% each of aliphatic and oxygenated (alcohols and ketones) hydrocarbons. The waste contains small amounts of chlorinated hydrocarbons and had a chlorine content of approximately 2%. The waste is air-atomized into the center of the pulverized coal nozzle. Freon-113 was added to the waste for these tests. The waste provided 40 to 60% of the heat input to the kiln during these test.

The flue gas was sampled for organic compounds by both VOST and MM5. Metals emissions were measured by atomic absorption analysis of the MM5 sample. The concentration of HCl in the stack gas was measured by sampling with a separate impinger train.

The waste feed contained significant quantities of four compounds that are listed in Appendix III: methylene chloride (MeCl_2), methyl ethyl ketone (MEK), toluene (MePh). Freon-113, which is not listed in Appendix VIII, was added to the waste fuel as a refractory surrogate. The methylene chloride data were considered unreliable because large quantities of this compound were found on the field blanks and because only 28% of the total quantity of this compound was found on the first of the two Tenax® tubes. These results are indicative of contamination. The MEK data were difficult to interpret. Sixty percent of the MEK was on the first tube but the emission rate during the baseline tests exceeded the emission rate during the co-fired test by 40%. If the co-fired emission rates are accepted as measured, the DRE of this compound was 99.99%. Subtraction of the baseline run emission rate from the co-fired run emission rate prior to calculating the DRE would produce a nonsense result. The toluene results were similar to the MEK results, i.e., 99% of the toluene was on the first tube but the baseline emission was equal to the co-fired emission. The toluene field blank results were all acceptably low so, contamination seems unlikely. The emission rate of toluene was unaffected by the addition of toluene to the fuel. The DRE of toluene (no adjustment for baseline emission rate) was 99.96%. The 1,1,1-trichloroethane results are somewhat better, though they too exhibit some signs of being influenced by contamination; on the average, 63% of the 1,1,1-trichloroethane was found on

the first tube but the baseline emission rate was 30% of the co-fired emission rate. The DRE of 1,1,1-trichloroethane, calculated without subtracting the baseline emission, was 99.991 to 99.999%. No Freon-113 was detected in any sample. Freon-113 was added to the waste only on test days 8 and 9. A significant concentration of Freon-113 was found in the sample of waste fuel taken on test day 7. The average DRE of Freon-113 for these three test days was greater than 99.9987%.

The results of these tests, summarized in Table 4.4, indicate that the kiln destroyed 99.99% of the hazardous material in the waste. This does not appear to be true of toluene. However, toluene emission rate during the co-fired tests was not higher than during the baseline tests.

Site C

Site C was a lightweight aggregate process capable of producing 20,000 pounds of expanded clay aggregate per hour. The kiln gases were exhausted through a pair of cyclone dust collectors and then through a cross flow water scrubber. Dust from the cyclones was recycled. The water from the spray scrubber was exhausted to a pond. Scrubber water was not recycled. The differences between this process and the cement process was that the maximum kiln temperature was only 2100°F (opposed to 2800° to 3000°F in cement kilns) and the product was not alkaline and thus did not scavenge HCl from the kiln gases. Note that although the maximum product temperature was 700°F less than in the cement process, the flame temperature was not less. Less fuel was burned per unit of product. The temperature of the flame was equal to the cement kiln flame temperature, but the length of time that the gases remain above 2800°F was shortened by the relatively high proportion of product in the kiln. The liquid waste was trucked directly to the site by generators. It consisted primarily of solvents, alcohols (10%), esters (5%), and aromatic compounds (15%). The waste fuel provided approximately 55% of the total heat input to the kiln during these tests. The balance of the fuel was provided by pulverized coal. The POHCs of the fuel and their approximate concentrations are given below:

<u>POHC</u>	<u>Weight Percent in Waste</u>
Methyl ethyl ketone (MEK)	2.5
Methyl isobutyl ketone (MIBK)	2.0
Tetrachloroethylene (Perc)	0.1
Toluene (MePhi)	8.3

The VOST was used for measurement of the emission rates of the four POHCs listed above. The VOST was modified, however, by insertion of an impinger upstream of the first Tenax® trap. This was done to alleviate the plugging of the Tenax® traps by condensate that caused abortion of most of the first six VOST runs. Unfortunately, the condensate was neither measured nor saved for analysis. A post hoc attempt was made to account for the quantity of POHCs that may have been lost in the condensate. The procedure consisted of estimating the volume of condensate (5 ml), and measuring the concentration of the POHCs in the scrubber effluent. The concentrations of POHCs in the condensate was assumed to equal the concentration in the scrubber effluent. This calculation probably underestimated the quantity of POHCs in the condensate since the scrubber effluent consists of condensate that has been diluted by scrubber water. Not accounting for the dilution may have caused

the POHC concentration in the condensate to be underestimated by approximately an order of magnitude. Since, the quantity of POHCs in 5 ml of the scrubber effluent was approximately equal to the quantity on the Tenax® tubes the emission rates may be underestimated by an order of magnitude.

The overall test average DREs for the various POHCs are given below. These values are those reported by the testing organization:

<u>POHC</u>	<u>DRE</u>
MEK	99.998
MIBK	99.998
Perc	99.997
Toluene	99.999

They include the estimates of the POHC concentrations in the condensate that were based on the scrubber effluent concentrations. As was noted, the estimated condensate concentrations may be low by a factor of 10, therefore, the DREs may be high by one significant figure (99.999 may be only 99.99). With this adjustment, the DREs of the POHCs approach or equal 99.99%. However, it is not possible to conclude that 99.99% was attained for certain during these tests.

Site D

Site D was a lime kiln with a capacity of 17,000 pounds of lime per hour. Limestone (CaCO_3) that was quarried nearby was heated to approximately 2000°F to drive off water and carbon dioxide. The product lime was sold as barn lime and hydrated lime, $\text{Ca}(\text{OH})_2$. Approximately 6.5 million Btu of heat per ton of product provided by a mixture of petroleum coke (90%) and natural gas (10%), was needed by the process. Secondary kiln air was drawn, preheated, from the lime cooler. The kiln exhaust gases passed through a series of large radiator coolers and then through a baghouse before being exhausted.

The waste fuel consisted primarily of lacquer thinner solvents, alcohols, still bottoms, paint wastes, and a smaller proportion (4%) of chlorinated hydrocarbons. The waste was spiked with perchloroethylene (perc) and trichloroethylene (TCE) to bring its chlorine concentration to 3%. In addition to these compounds, four other hazardous constituents were present in the waste at concentrations sufficient to allow computation of their DREs. They were: methylene chloride (MeCl_2), methyl ethyl ketone (MEK), 1,1,1-trichloroethylene (111-TCE), and toluene. The waste fuel contributed from 8 to 36% of the heat input to the kiln during these tests. Waste was atomized from a 1-inch diameter pipe that was strapped to the gas burner. The coke was fed through the annular space surrounding the gas and waste supply pipes.

The emission rates of the two POHCs, perc and TCE, were measured by the VOST. HCl was determined by specific ion electrode measurement of the chloride in the impingers of the Method 5 train.

Typically, the fraction of the various POHCs that were trapped on the first Tenax® tube ranged from about 50 to 80%. A few pair of tubes contained less than 50% of one or more POHCs on the first tube. The quantity of the various POHCs found on the field and laboratory blanks ranged from 50 to 200%

of the quantity found on the associated sample tubes. These things appear to be the result of contamination. On the other hand, some 20% of the VOST results are based upon analysis of only one of the two tubes, either the Tenax® or the Tenax®/charcoal tube having been lost (by breakage or GC/MS malfunction). The total emission rates (Table 4.4) were low and the DREs high, however. The baseline (no waste feed) emission rates were quite near the co-fired emission rates. In fact, the difference between the baseline emission rates and the co-fired emission rates were statistically significant for only toluene and methylene chloride. Thus, it is concluded that the kiln destroyed at least 99.99% of the hazardous organic constituents in the waste.

Site E

Site E was a dry process cement plant with a capacity of 120,000 pounds of cement per hour. This plant had an unusually long (520 feet) kiln with an unusually long (10 seconds) gas retention time. The added length, and thus retention time, was needed because the kiln did not have a raw materials preheater. The kiln gases were exhausted through cyclones and then through an ESP. The dust collected in the cyclones and in the first three ESP banks was recycled. Dust collected in the fourth ESP bank was considered waste.

The waste burned during these tests was the still bottoms waste of a commercial solvent reclaimer that normally was used as supplemental fuel by another cement plant. This waste contained approximately 11% toluene, 8% methyl ethyl ketone (MEK), 4% methylene chloride (MeCl_2), and 2% 1,1,1-trichloroethane. The balance was alcohols, esters, and alkylated aromatic compounds. The four compounds listed were taken as the POHCs for these tests. The fuel had a chlorine content of 4.5% and a gross heating value of approximately 12,300 Btu/pound. The waste was air-atomized into the primary fuel (coal and petroleum coke) flame. The waste provided an average of 13% of the heat input to the kiln during these tests.

The stack gas concentrations of the POHCs were measured by M23. The lowest concentrations of the POHCs that were detectable by M23 were higher than the concentrations of the POHCs in the stack gas. Thus, even though the flow rates of the various POHCs into the kiln were appreciable it was not possible to measure them in the stack gas at 99.99% DRE. No POHC was detected during the co-fired runs but the maximum calculable DRE values ranged from 99.73 to 99.97%. The data demonstrate that the DREs were at least that good but the method used could not demonstrate 99.99% DRE.

Site F

Site F was a dry process cement plant with a production capacity of 120,000 pounds of clinker per hour. This kiln was similar to Site E in that there was no raw material preheater so the kiln was unusually long (520 feet) with an unusually long gas retention time. The kiln was normally fired with a mixture of pulverized coal and coke. Approximately 65% of the hot air from the clinker cooler was used as preheated combustion air, the balance was discharged through a cyclone and ESP. The kiln exhaust gases were cooled by water sprays to 700°F and then passed through a cyclone and an ESP for particle removal. Approximately 90% of the 225 tons per day of dust that was collected was recycled. The remaining 25 tons per day was wasted.

A waste fuel supply pipe was attached to the underside of the coal/coke burner. The waste was pressure atomized. The waste itself was composed of hydrocarbon solvents and resins and solids that are typical of paint wastes. The waste was spiked with 1,1,2-trichloro-1,2,2-trifluoroethane (Freon-113). This compound served as a surrogate for refractory organic compounds. The other POHCs that were designated (listed below) were constituents of the waste.

<u>Compound</u>	<u>Weight Percent in Waste Fuel</u>
Freon-113	0.8
Methylene chloride (MeCl ₂)	0.4
Methyl ethyl ketone (MEK)	0.9
1,1,1-trichloroethane	1.0
Toluene	21.6

Two baseline (no waste fuel burned) runs were done. These were followed by three co-fired runs. The percent fuel replacement during these runs was 25, 37, and 42% for a test average of 35%. The POHC emission rate data from the final run (42% fuel replacement) were invalidated by the analytical laboratory because the GC/MS internal standard responses were outside of the quality control range.

The emission rates of POHCs and other organic compounds were measured by both the VOST and MM5 procedures. DRE calculations were based exclusively on the VOST analyses. MM5 samples were analyzed by GC/MS for higher boiling fuel constituents and for PICs. The VOST sample results were difficult to interpret. There was evidence that the samples became contaminated. There were two prime indicators of contamination: high blanks, especially methylene chloride, and the baseline emission rates (see Table 4.4) were higher for all compounds than the co-fired emission rates. Thus, contamination may have contributed significantly to the measured emission rates.

The data (Table 4.4) show the DRE of all compounds, except methylene chloride, to be well in excess of 99.99%. The methylene chloride data were discounted because of the potential for contamination. There was no indication (other than the MeCl₂ data) that the kiln did not achieve 99.99% DRE of the hazardous compounds in the waste fuel. The MM5 data, which were less prone to be influenced by contamination because of the larger quantities of compounds in the samples, indicated that the DRE of other waste fuel constituents exceeded 99.99%. It is most likely that the MeCl₂ data are in error.

Site G

The kiln tested at Site G was a wet process unit that was producing 70,000 pounds of clinker per hour. The kiln exhaust gases were cooled and then passed through a baghouse for particle removal prior to their discharge to the stack. The test report did not state how much of the collected dust was recycled nor did it describe whether the clinker cooler exhaust was used as kiln combustion air. The primary fuel burned at the plant was No. 6 fuel oil.

The waste was a mixture of solvents, primarily alkanes and esters. The chlorine content of the waste averaged 25% (range 6 to 35%). Most of the

chlorine in the waste was contributed by three compounds: methylene chloride (MeCl_2), chloroform (MeCl_3), and carbon tetrachloride (MeCl_4). These three compounds were designated POHCs. The waste fuel burner pipe was located along the inside of the annular primary combustion air conduit of the oil burner at the 2 o'clock position. The waste was sprayed under pressure into the oil flame but it was not air or steam atomized. The waste fuel provided approximately 10% (range 3 to 14%) of the heat input into the kiln during these tests.

The data from this site (Table 4.4) do not appear to be similar to the data from the other six sites. For example, the MeCl_4 emission rate measured during the baseline run is approximately 200 times higher than the average POHC emission rate during the co-fired runs at the other six sites. The baseline results for the other two POHCs, while not so extreme, are also high. The measured co-fired emission rates are even higher. Note that the baseline emission rates have been subtracted, the co-fired emission rates given in Table 4.4 are the measured rates less the baseline rates. The DREs were calculated using the baseline adjusted-emission rates. Apparently, this kiln did a relatively poor job of destroying hazardous compounds.

Particle Emissions

The rates of emission of particles during the baseline and co-fired tests at the various sites are given in Table 4.5. Combustion of hazardous waste had an effect on the emission of dust at only Site F. The test report states that the chlorine from the waste fuel appeared to concentrate in the small particles. The observation was made at several sites. The Site F operators reported that the added chloride changed the resistivity of the cement dust enough to seriously degrade the performance of the ESP. The operators were apparently aware that this might occur and had already consulted with the manufacturer of the precipitator to explore possible corrective actions. The increased emissions at Site G were the results of the breaking of several bags in the fabric filter. The problem first appeared during one of the baseline runs and was unrelated to the firing of hazardous waste. The particle emission rates for the six co-fired runs performed after replacement of the broken bags was 19.5 pounds per hour (0.068 grains per dry standard cubic foot at 7% oxygen).

Sites F and C appear to be the only two that did not comply with the particle emission rate standard for hazardous waste incinerators of 0.08 grains per dry standard cubic foot, corrected to 7% oxygen. Site F, as was mentioned, apparently developed a correctable dust resistivity problem as a result of burning the chlorinated waste. Site C was equipped with only a cross flow water scrubber that cannot be expected to remove particles efficiently.

HCl Emissions

The measured rates of chlorine input in the fuel (waste fuel plus primary fuel) were compared to the stack gas emission rates of gaseous chloride in Table 4.6. These data demonstrate that the alkaline product effectively scavenges HCl from the combustion gases. Several of the test reports noted that the bulk of the chloride appeared to be associated with the fine particle fraction of the dust that was suspended in the kiln exhaust gases. The

TABLE 4.5
PARTICLE EMISSION RATE FROM KILNS

Site	Particle Emissions			
	Baseline		Co-Fired	
	lb/hr	gr/dscf ^a	lb/hr	gr/dscf ^a
A	NM ^b	NM	NM	NM
B	19.7	0.053	19.0	0.052
C	c	c	11.7	0.182
D	2.0	0.013	2.2	0.013
E	49.6	0.069	58.4	0.080
F	52.5	0.164	240.0	0.828
G	21.7	0.069	30.7	0.117
G ^d			19.5	0.068

^a Corrected to 7% O₂ as required by incinerator regulations.

^b Not measured.

^c No baseline run at this site.

^d Co-fired runs after replacement of broken bags.

TABLE 4.6
HYDROCHLORIC ACID EMISSION RATE FROM KILNS

Site	Chlorine in Fuel (lb/hr)	Chloride in Stack Gas (lb/hr)	Removal Efficiency
A			
Baseline	---	0.55	-----
Co-Fire	156	1.03	99.34
B			
Baseline	13.3	<1.8	86.47
Co-Fire	129.3	3.52	97.28
C			
Baseline	a	a	-----
Co-Fire	22.0	0.051	97.77
D			
Baseline	1.6	0.20	87.50
Co-Fire	28.6	0.44	98.46
E			
Baseline	36	3.7	89.72
Co-Fire	178	3.2	98.20
F			
Baseline	22.9	2.9	87.34
Co-Fire	149.1	25.3	83.03
G			
Baseline	ND ^b	0.34	-----
Co-Fire	351.0	0.96	99.73

a No baseline run at this site.

b No chlorine detected in No. 6 oil.

rate of waste of dust collector fines had to be increased to prevent excessive buildup of chloride in the product at the cement plant. The increased rate of production of waste dust was not seen to be a major hardship. The rate of emission of HCl at Site F was significantly higher during the co-fired tests than at the other sites. It is believed that these samples were contaminated by the high concentration of chloride laden dust. Sites E and F were similar facilities and showed similar HCl removal on baseline tests. Site C was an expanded clay aggregate kiln. Neither the raw material nor the product of this plant was sufficiently alkaline to absorb significant amounts of chloride. However, the kiln was equipped with a wet scrubber that efficiently removed the HCl from the gas stream. If Site F is ignored, the kiln dust clinker and control equipment collectively removed 97% or greater of the HCl from hazardous waste combustion on six kiln tests.

Products of Incomplete Combustion (PICs)

Compounds not present in the fuel or waste fuel burned appeared in the exhaust gas of several sources. There are two general mechanisms by which PICs may be formed. One is formation of products of partial oxidation of fuel components. The formation of aldehydes during incineration of refuse is an example of this mechanism. Formation of low molecular weight hydrocarbons (C_1 - C_6) during combustion of coal and heavy oil fuels is another example. The second general mechanism is a reaction of free radicals or other molecular fragments produced in the high temperature flame to produce different compounds. The products of these reactions may have higher molecular weights than the fuel components. Benzene, polynuclear aromatic compounds, and soot particles appear to be formed in this manner.

There are also means by which fuel-absent compounds may appear in the exhaust gas that are unrelated to the combustion. Perhaps the major source of fuel-absent compounds is the raw materials. These materials are sedimentary, usually rock, but sometimes oyster shells, argonite, etc., all of which inherently contain greater or lesser degrees of biogenic material, often residual. Levels can be as high as a few percent of the total raw materials. This biogenic material contain organic compounds that often will be driven off at low temperatures and appear in the exhaust gases without having any opportunity to be combusted. Other sources of fuel-absent compounds include evaporation of lubricating oils from mechanical equipment downstream of the kiln and compounds that were present in the ambient air that was used in the furnace. The water used in wet process cement slurry or in the scrubbers at lime and aggregate kilns may be a source of organic compounds. These may exist in the fresh water supply, be introduced by other processes where water re-use is practiced, or they may have been added in the form of fungicides or other water treatment chemicals. Compounds from these sources are not PICs, but rather, are flue gas contaminants.

A third source of fuel-absent compounds in samples taken from the flue gas is contamination of those samples during sampling or analysis. This contamination can be external. For example, large amounts of Freon were identified in Tenax®-GC samples from one site. Perhaps the refrigerator in which the samples were stored was defective and Freon invaded the samples. The contamination can also be internal. Several investigators have reported that benzene appears to be a product of thermal degradation of Tenax®-GC. Others

have attributed the presence of several C₅-C₉ alcohols and ketones in exhaust gas samples to the degradation of XAD resin. In addition, there is evidence that compounds sorbed by porous polymer resins are not quantitatively removed during their preliminary cleanup. Successive cleanings release additional amounts of these compounds. Thus, it is possible that a measured compound may be the residue of some past sample or contamination.

Benzene emissions were reported at Sites A and B. These emissions (Table 4.7) occurred during both the baseline and the co-fired test runs. There was not enough data to establish whether these reported emissions were due to primary fuel combustion, waste fuel combustion or analytical artifacts. No other significant PIC emissions were reported.

Metals Emissions

The metals emissions data that were available (Table 4.8) did not indicate that the combustion of waste fuel increases the rate of emission of metals by more than would be expected from their increased firing rate. For example, the firing rate of lead (Pb) was five times higher during the cofired run at Site B than it was during the baseline run, and its emission rate increased by a factor of five. The fraction of those metals in the fuels that have been shown to concentrate in small particles (all except Ni in Table 4.8) emitted to the atmosphere was higher than the fractional emission of other metals. The amount of even the volatile elements emitted was less than the amount in the waste fuel; i.e., approximately 15% of the mercury (Hg) in the waste fuel at Site B was emitted.

None of the metals sought at Site A were detectable in the stack gas particles, primarily because the total mass of sample was low. The waste burned at Site G contained negligible concentrations of metals. Thus, no baseline to co-fired test comparisons were possible for these sites. The mass emission rates of all dust constituents was high during the co-fired tests at Site F because the chloride increase in the dust that occurred from burning the waste changed the dust resistivity and degraded the performance of the precipitator. Those emissions were not representative of normal operation and were not included in the table.

The metals emissions data at Site C demonstrated the effect of concentration of certain elements in small particles. This site was equipped with a crossflow water scrubber that was ineffective at removal of small particles. The fraction of the fuel-contained volatile metals (Pb and Cr) emitted in the stack gas was approximately four times higher at this site than it was at the other sites.

Combustion Gases and DRE

Table 4.9 contains average values for combustion gases plus NO₂ and the DREs of two hazardous compounds. No single hazardous compound was common to all sites. The CO₂ concentrations were high and, except at Site C, were not representative of combustion conditions, since at all other sites limestone was being calcined. There was no indication why both the unburned hydrocarbons and the NO_x concentrations at Site E should be so much higher than they were at the other sites. Site E DRE data cannot be compared to the other sites since there was no upper bound for them. No meaningful correlation

TABLE 4.7

PRODUCTS OF INCOMPLETE COMBUSTION FROM KILNS

Site	Compound	Baseline (lb/hr)	Co-Fire (lb/hr)
A	benzene	0.4	0.6
	PCBs	<0.001	<0.001
B	perchloroethylene	0.007	0.008
	benzene	0.29	1.25
	chloroform	0.001	0.005
C	NO ATTEMPT MADE TO MEASURE PICs AT THIS SITE		
D	NO ATTEMPT MADE TO MEASURE PICs AT THIS SITE		
E	NO ATTEMPT MADE TO MEASURE PICs AT THIS SITE		
F	NO COMPOUNDS NOT IN WASTE WERE FOUND IN STACK		
G	aromatic compounds	----	0.0002 ^a
	polynuclear aro-		
	matic compounds	----	0.0002 ^a
	sulfur	----	0.00003 ^a

^a Compounds found during co-fired runs that were not found during baseline run.

TABLE 4.8
METALS EMISSION RATES FROM KILNS

Site	Metal	Mass Flow Rates (lb/hr)			
		Baseline Tests		Co-fired Tests	
		Fuel Total	Exhaust Gas	Fuel Total	Exhaust Gas
A	None Detected				
B	Cd	0.007	<0.00001	0.13	0.005
	Cu	<0.345	<0.0004	0.676	<0.001
	Hg	<0.002	<0.0001	0.009	0.001
	Ni	0.809	0.001	2.113	<0.0004
	Pb	0.485	0.012	2.396	0.055
	Se	0.009	0.004	0.080	0.027
C	Cd			0.034	0.006
	Cr			0.80	0.016
	Pb			5.0	0.43
D	Pb	0.098	<0.006	0.265	0.004
	Zn	0.032	0.001	0.221	0.005
E	Mass flow rates of metals were not calculated				
F	Electrostatic precipitator malfunction				
G	Metals in waste fuel were negligible				

TABLE 4.9

AVERAGE FLUE GAS CONCENTRATIONS OF COMBUSTION GASES FROM KILNS

Site	O ₂ (%)		CO ₂ (%)		CO (ppm)		TUHC ^e (ppm)		NO _x (ppm)		DRE (%)	
	Base- line	Co- Fire	Base- line	Co- Fire	Base- line	Co- Fire	Base- line	Co- Fire	Base- line	Co- Fire	Methylene Chloride	Toluene
A	8.3	8.9	17.0	16.2	64	40	a	a	682	486	99.998	a
B	12.3	12.0	12.9	13.1	212	190	10.2	21.0	371	477	a	99.990
C	b	13.6	b	1.5	b	a	a	a	b	161	a	99.998
D	7.1	5.6	23.6	22.0	426	646	8.4	3.6	376	422	99.997	99.999
E	5.3	5.4	24.2	23.3	3.3	4.0	1255 ^c	469 ^c	1600	959	>99.909 ^d	>99.968 ^d
F	6.5	7.2	23.6	22.4	37.5	38.7	2.5	5	620	814	99.960	99.995
G	12.1	11.8	15.3	14.7	166	326	9.0	12.7	136	68	99.662	a

a Not measured.

b No baseline run at this site.

c Total gaseous nonmethane organic matter.

d None detected. DRE calculated based on minimum detection limit of Method 23.

e Total unburned hydrocarbons.

between DRE and any PIC could be found. Nor was there any indication that a threshold concentration of any combustion gas, beyond which DRE was attenuated, exists.

Regulatory Implications of Test Burn Data

The data indicate that, where alkaline material is produced, the HCl that results from combustion of chlorinated organic material is absorbed by the material. Thus, the HCl emissions are all low. There is no indication that the burning of hazardous waste affected the rate of particle emissions, except at Site F, where the excess chloride in the dust during co-firing appeared to change the resistivity of the dust and degrade the performance of the ESP. This was not observed at the other three sites (B, G, and E) that were equipped with ESPs and was thought to be correctable at Site F. Only Site C, which was equipped with only a crossflow water scrubber, emitted particles at a rate in excess of 0.08 grains per dry standard cubic foot, corrected to 7% oxygen. Thus, it is evident that hazardous materials can be burned in kilns without causing excess emissions of HCl and particles.

The metals emissions data indicated the need for a fuel specification/control device specification regulation. The combination of a control device that is ineffective for small particle removal and a high concentration of volatile, hazardous metals could result in excessive emissions.

The DRE data demonstrated that it is possible for kilns to achieve adequate (99.99%) destruction of hazardous organic compounds. However, it was not demonstrated conclusively at all sites nor have a complete spectrum of all equipment types and operating conditions been sampled. Kilns are subject to variations in design and operating conditions in response to variations in raw material compositions and product specifications. The population of kilns cannot be easily categorized. Therefore, to be environmentally certain, regulations would need to require that a trial burn be done at each site where combustion of hazardous organic compounds was contemplated. Care should be exercised in the design of these trial burns to assure that the combination of POHC selection and stack gas test method will bound the DRE at 99.99% or higher.

EVALUATION OF TEST BURN RESULTS FOR ASPHALT AGGREGATE KILNS

Interest in the level of emissions from burning waste derived fuels in asphalt concrete plants has grown as the practice has increased throughout the industry. The State of Texas conducted emission testing at an asphalt concrete plant (Reference 13) due to high occurrence of chlorinated compounds in their fuel. The results showed that destruction efficiencies did exceed 99.99% for most chlorinated species tested. The report concluded that downwind concentrations neither adversely affect the environment nor posed any health effects from public exposure.

Sampling performed for the Massachusetts Asphalt Pavement Association (MAPA) on spiked waste derived fuel (Reference 14) showed that the particulate limits (NSPS) were met but trichloroethylene had a destruction removal efficiency of 99.75% and hydrochloric acid was emitted at 4.81 pounds per hour with an overall removal efficiency of 81.96%.

EPA tested three asphalt plants to provide data to support developing regulations for burning hazardous waste in asphalt plants (Reference 15). The goals for the test program were to:

- o Determine the DRE of two POHCs (perchloroethylene and chlorobenzene) in a batch plant and a drum-mix plant.
- o Determine metals removal efficiency by a baghouse and a scrubber.
- o Evaluate HCl emissions from a baghouse-equipped unit and a scrubber-equipped unit.

Initially, it was decided to sample a drum-mix plant equipped with a baghouse (Plant A) and a batch plant equipped with a scrubber (Plant B). When Plant B was tested, cyclonic flow in the stack precluded sampling for particulate (metals). A third plant (Plant C) was added to the program to supplement the metal data for the "batch plant equipped with a scrubber".

Plant A is a 1984 vintage Barber-Greene 400 ton per hour drum-mix plant equipped with a baghouse. The unit is described as a "six-pack" because it is mounted on six flat bed trailers for easy relocation. The fuel was 100% recycled oil, primarily from industrial sources. Due to heavy rain prior to the test program, the aggregate had a high moisture content. Therefore, production was limited to 295 tons per hour during the test program. The average fuel consumption was 500 gallons or 100% waste oil per hour or about 1.7 gallons per ton of concrete produced.

Plant B is a 1950's vintage batch plant equipped with primary cyclone and low efficiency scrubber. The fuel oil was 100% recycled oil, primarily crankcase oil. The hourly production rate was 195 tons per hour burning 390 gallons per hour of 100% waste oil. Fuel consumption was 2.0 gallons per ton of product.

Plant C is a 1950's vintage batch plant equipped with a primary cyclone and a medium to high efficiency scrubber. The fuel fired was 100% recycled crankcase oil. Plant C had the lowest production capacity at 76 tons per hour, burning 165 gallons per hour of 100% waste oil. This is equivalent to 2.2 gallons per ton of product.

Plant A represents a typical modern type plant (drum-mix design) with a high efficiency particulate collector (baghouse). Plants B and C represent older type plants (batch process) with low to medium efficiency particulate collectors (water scrubbers).

Two exhaust gas test locations were sampled on the plants. The inlet of the control device was sampled for particulates and metals only. The stack was also sampled for particulates and metals to allow determination of control equipment removal efficiencies. Additionally, the exhaust stack was sampled using EPA MM5 for POHCs and a MM6 for HCl. CEMs were used for monitoring oxygen, carbon monoxide, and total hydrocarbons. The CEMs were used primarily for monitoring the process to insure against sampling during upset conditions.

Table 4.10 summarizes the production data for the three units and contains stack exhaust gas data including moisture, temperature, oxygen content, and gas flow rate at standard conditions.

Two target compounds were used to measure the DRE of the plants; perchloroethylene and chlorobenzene. The compounds were injected into the burner fuel line between the fuel tank and the burner. This approach provided several advantages over spiking the fuel in the tank.

- o The fuel tanks were not contaminated.
- o It was not necessary to mix the fuel to avoid stratification.
- o It was much easier (and more accurate) to monitor the actual POHC, injection rate using calibrated rotometers.

The sampling method used was EPA MM5 with XAD-2 as the sorbent trap. The analytical method followed SW846 guidelines for GC/MS. The two target compounds were injected into the fuel at rates to approximately 1% by weight of the fuel. Samples of the fuel oil were collected during each test and submitted for analysis. The actual feed rates and feed oil concentrations of the target POHCs are provided in Table 4.11.

The measured DREs are presented in Table 4.12. Plant A is the drum-mix plant and Plant B is the batch plant. In most cases the DREs are reported to one place beyond actual sensitivity of the method. That estimated decimal is underlined to indicate that it is an approximation. In a few cases, perchloroethylene was observed but it was below the reported detectable limit and outside the range of calibration. Those observations are preceded by an "approximate" sign. All of the other samples contained no detectable amounts of perchloroethylene. The DREs were calculated using the minimum detectable laboratory values and are preceded by a "greater than" notation.

The drum-mix plant demonstrated 99.99% DRE or better for both POHCs for all tests. The batch plant results provided only estimates for the fourth decimal of the DRE percent of perc. Chlorobenzene was measurable in all of the batch plant sample and was found to be in excess of the minimum amount showing 99.99% DRE. The average DRE for chlorobenzene was 99.95% in the batch plant for the three tests.

Although a single series of tests on one of each type of plant does not provide conclusive evidence, the results do indicate that asphalt plants may be effective and environmentally sound incinerators of waste oil and some hazardous wastes. The DRE of POHCs can be as high as 99.999% under optimum conditions but may be as low as 99.95% under less optimum conditions.

Metals were measured in samples of the asphalt cement, aggregate feed, waste fuel oil, and the control device inlet and outlet particulate catch. Table 4.13 shows the concentration of the metals in the aggregate and fuel oil at Plants A and C. Plant B was sampled for organics and HCl only, due to cyclonic flow in the stack. The third process feed material, asphalt cement, was also analyzed for metals at Plant A because it represented a potential source for metal emissions. All of the metals were below the detectable level in the asphalt cement samples.

TABLE 4.10

AVERAGE PRODUCTION DATA AND EXHAUST GAS CHARACTERISTICS

Plant	Type	Controls	Production Data		Exhaust Data			
			Waste Oil (gph)	Asphalt Concrete (tph)	Gas Moist. (%)	Gas Temp. (°F)	Oxygen Content (%)	Gas Flow (dscfm)
A	Drum- mix	Baghouse	501	295	37	280	10.1	21,900
B	Batch	Spray scrubber	389	195	18	142	12.6	24,460
C	Batch	Venturi scrubber	165	76	12	134	16.0	16,470

TABLE 4.11

POHC FEED RATES AND RESULTING CONCENTRATION OF THE FUEL

	Perchloroethylene		Chlorobenzene	
	Grams/Minute	Percent of Fuel	Grams/Minute	Percent of Fuel
PLANT A				
Test 1	250	0.87	255	0.88
Test 2	178	0.66	269	1.00
Test 3	406	1.40	319	1.10
Test 4	435	1.50	377	1.30
PLANT B				
Test 1	85	0.34	185	0.74
Test 2	86	0.44	176	0.90
Test 3	86	0.44	176	0.90

TABLE 4.12

MEASURED DRES AT PLANTS A AND B

Perchloroethylene (%)		Chlorobenzene (%)
PLANT A		
Test 1	= 99.997	99.991
Test 2	= 99.999	99.988
Test 3	> 99.997	99.9915
Test 4	> 99.9997	99.9941
PLANT B		
Test 1	> 99.95	99.94
Test 2	= 99.99	99.95
Test 3	> 99.92	99.96

TABLE 4.13

METALS CONCENTRATIONS IN THE AGGREGATE AND
WASTE OILS OF PLANTS A AND C

Metal	Plant A		Plant C	
	Aggregate (ppm)	Waste Fuel (ppm)	Aggregate (ppm)	Waste Oil (ppm)
Arsenic	not analyzed	---	---	---
Aluminum	1,300	295	5	0-1
Boron	<1.0	<1.0	1,930	22
Barium	4.4	23.8	10	26
Cadmium	4.7	<0.3	0-1	1
Calcium	150,000	42.5	58,300	571
Chromium	2.7	287.5	11	3
Iron	5,900	570	8,120	188
Lead	11	71.8	7	416
Magnesium	48,000	11.5-14	3,590	286
Manganese	380	5.3	88	7
Molybdenum	5.1	8.1-9.6	4	7
Nickel	8.0	3.4	7	4
Silicon	270	69.3	665	150
Silver	<3.0	1.7-3.9	0-1	0-1
Sodium	160	0-20	23	385
Tin	<30	0-50	0-10	0-10
Titanium	42	1.8-2.3	73	0-5
Vanadium	7.3	10	10	12
Zinc	15	67	10	593

Although a few of the metals were present in the oil in much greater concentrations than in the aggregate (Plant A - barium, chromium, lead, and zinc; Plant B - boron, barium, lead, sodium, and zinc), the feed rate of the aggregate was over 100 times the feed rate of the fuel oil on a weight to weight basis. This precluded tracking the fate of a given specific metal from the oil through the system.

Another approach was taken for evaluating the metals data. Based upon the relative feed rate of the aggregate and fuels oils (hundreds of tons per hour of aggregate to 1 or 2 tons per hour of oil) it was hypothesized that the metal concentration of the particulate catch at the control device inlet should be similar to the metal composition of the aggregate. Theoretical metal loadings were calculated in pounds per hour at the baghouse inlet assuming that the particulate catch there was predominantly aggregate. The potential contribution from the waste oil was then calculated assuming that all of the metals in the oil reached the control device inlet. Finally, the actual metal loading was determined from analyses of the Method 5 particulate catch. The results are compared in Tables 4.14 and 4.15 for Plants A and C, respectively. From these results it appears that the waste oil contributes a significant portion of aluminum, barium, chromium, iron, lead, molybdenum, nickel, silicon, and zinc to the particulate material at the control device inlet of Plant A, and boron, barium, chromium, lead, sodium, and zinc at Plant C. The results are not conclusive, but are indicative especially that the barium, chromium, lead, and zinc from the waste oils are emitted from the asphalt kilns into the control equipment.

Table 4.16 shows the overall particulate control efficiencies of the baghouse (99.75% efficiency) and scrubber (99.24% efficiency). Table 4.17 presents the range of measured control efficiencies for each of the metals at the two plants. The efficiencies for the metals deduced as being from the fuel oils are underlined in the table. Most metals were controlled as efficiently as the total particulate loading in the baghouse. However, the scrubber was not as efficient for several metals including lead, sodium, and zinc. A baghouse is generally as efficient for small particulate as it is for large, whereas the efficiency of the scrubber decreases with the particle size. Some of the metals are possibly more concentrated in the finer particles and therefore collected more efficiently by a baghouse than a scrubber.

Plants A, B, and C were all sampled for HCl emissions using a midjet impinger train. The samples were analyzed by specific ion electrode. The inlet of the venturi scrubber on Plant C was also tested. Samples of the feed oils were also analyzed for total chloride content. The chloride concentrations in the waste oils burned at the three plants varied significantly. The concentrations were 1.6, 0.5, and 0.1% by weight for Plants A, B, and C, respectively.

Plant A burned industrial waste oil which contains "extreme pressure" (EP) lubricants which are long chain chlorinated hydrocarbons.

The stack gas samples at all locations were found by analysis to be below the detection level of the method, which is approximately equivalent to 30 ppm of HCl in the stack gas or less than 0.5 pounds per hour at Plants A and B. The sample volume was increased at Site C so that the minimum detectable concentration was 0.4 ppm or equivalent to 0.03 pounds per hour. HCl

TABLE 4.14

COMPARISON OF THEORETICAL METAL CONTRIBUTIONS FROM
AGGREGATE AND WASTE OIL TO ACTUAL MEASURED METAL
LOADINGS AT THE CONTROL DEVICE INLET, PLANT A

Metal	Aggregate*	Waste Oil	Aggregate Plus Waste Oil	Actual Measured Loading (pounds per hour $\times 10^{-2}$)
Aluminum	200	113	313	374
Boron	<0.15	<0.38	<.38	ND
Barium	0.68	9.08	9.76	7.79
Cadmium	0.72	<0.11	0.72	0.2
Calcium	23,000	16.3	23,000	25,400
Chromium	0.41	110	110	101
Iron	906	218	1,124	1,165
Lead	1.7	27.4	29.1	39.8
Magnesium	7,370	4.88	7,370	9,330
Manganese	58.3	2.01	60.3	38.2
Molybdenum	0.78	3.38	0.78	3.75
Nickel	1.23	1.28	2.51	3.68
Silicon	41.5	26.5	68.0	68.6
Silver	<0.46	1.0	<1.5	0.6
Sodium	24.6	<7.7	24.6	47.9
Tin	<4.61	<19.1	<19.1	9.63
Titanium	6.45	0.76	6.45	12.8
Vanadium	1.12	<3.83	1.12	1.68
Zinc	2.30	25.6	27.9	24.2

* Metal concentration in the bulk aggregate multiplied by the mass flow rate of dust in the flue gas.

TABLE 4.15

COMPARISON OF THEORETICAL METAL CONTRIBUTIONS FROM
AGGREGATE AND WASTE OIL TO ACTUAL MEASURED METAL
LOADINGS AT THE CONTROL DEVICE INLET, PLANT C

Metal	Aggregate*	Waste Oil	Aggregate Plus Waste Oil	Actual Measured Loading (pounds per hour $\times 10^{-2}$)
Arsenic	0.13	<0.12	0.13	0.23
Aluminum	52.4	2.72	55.1	150
Boron	<0.05	3.22	3.27	0.14
Barium	0.27	3.22	3.49	0.92
Cadmium	<0.03	0.14	0.17	0.04
Calcium	1,585	70.6	1,656	5,476
Chromium	0.30	0.32	0.62	0.97
Iron	220	23.3	243	445
Lead	0.20	51.5	51.7	13.48
Magnesium	97.6	35.4	133	371
Manganese	2.39	0.84	3.23	7.59
Molybdenum	0.10	0.90	1.00	0.36
Nickel	0.20	0.43	0.63	2.12
Silicon	18.1	18.6	36.7	10.7
Silver	<0.03	<0.12	0.15	0.05
Sodium	0.63	47.6	48.2	14.7
Tin	<0.27	<1.27	1.51	0.24
Titanium	1.98	<0.62	2.60	3.96
Vanadium	0.27	1.48	1.75	0.85
Zinc	0.27	73.4	73.7	21.0

* Metal concentration in the bulk aggregate multiplied by the mass flow rate of dust in the flue gas.

TABLE 4.16

PARTICULATE LOADINGS AND CONTROL DEVICE EFFICIENCIES

Plant	Sample Location	Particulate Concentration (gr/dscf)	Particulate Loading (lb/hr)	Control Efficiencies (%)
A	Inlet of Baghouse	8.9	1536	99.75
	Outlet of Baghouse	0.020	3.96	
C	Inlet of Scrubber	2.32	271.9	99.24*
	Outlet of Scrubber	0.015	2.07	

* The venturi was operating at a pressure drop of 16-17.5 inches of water.

TABLE 4.17

METAL CONTROL EFFICIENCIES: PLANT A AND C

Metal	Air Pollution Control Device Efficiencies	
	Plant A (%)	Plant C (%)
Arsenic	Not analyzed	87-100
Aluminum	100	100
Boron	NA	77-100
Barium	0-100	99-100
Cadmium	99-100	56-81
Calcium	98-100	100
Chromium	100	85-93
Iron	100	99
Lead	89-100	25-42
Magnesium	70-100	100
Manganese	98-100	97-99
Molybdenum	100	67-78
Nickel	0-100	94
Silicon	0-100	64-79
Silver	89-100	65-81
Sodium	10-100	36-56
Tin	0-100	0-36
Titanium	82-98	99-100
Vanadium	0-100	81-94
Zinc	35-97	25-28

was still not detected in any of the samples. The results are presented in Table 4.18 with the estimated equivalent HCl fuel feed rate.

HCl samples were collected at the inlet to the scrubber at Plant C. Because the results at the inlet of the Plant C scrubber were also below the detectable level (0.03 pounds per hour) it appears that the chlorides may be attracted to the aggregate which was 60% dolomitic lime at the sites tested.

It appears from these test results that burning chlorinated waste at chloride inputs at the levels used in the test are not likely to result in an air pollution problem. It is quite possible that the chlorides react with the calcium in the aggregate during the drying process. This has been demonstrated to occur during hazardous waste incineration in cement kilns.

EVALUATION OF TEST BURN RESULTS FOR BLAST FURNACE

A blast furnace should obtain greater than 99.99% DRE of toxic organics. However, no actual testing data are currently available to substantiate this expected performance. The single test conducted by EPA to evaluate typical waste disposal performance of a blast furnace when burning a liquid organic waste (Reference 16) is deemed invalid. There were several problems with the test. One major problem was that the waste feed POHC composition varied significantly during the test runs. This introduced error in the calculation of the DRE because of uncertainty in the amount of the POHCs in the feed. Another problem was that the POHC loading on some of the blanks for the VOST analyses were higher than the sample loadings. Also, for some of the POHCs, the DRE across the blast furnace was higher than that measured across the blast furnace and stoves. Still another shortcoming of the tests was that there were no provisions made for determining whether the POHC collected from the gas stream was PICs from the combustion of coke and fuel oil or was a constituent of the hazardous waste.

Although the single test was invalid, it is deemed very likely that POHCs would be destroyed by a blast furnace system with a DRE of over 99.99% because the fraction escaping destruction in the blast furnace is passed through another combustion device. It is expected that the stoves and boilers in which the off-gas is combusted should be capable of the same level of DRE as the industrial boilers for which testing has demonstrated over 99.99% DRE of toxic organics with regard to the off-gas that is flared, results from several studies (References 17-19) indicated over 98% DRE of the organic compounds can be expected if the flares are operated such that a stable flame is maintained, i.e., flame stays lit. Under unstable conditions, however, the flare destruction efficiency can be reduced to approximately 55% (Reference 17).

To date, EPA has not conducted any tests to determine the fate of metal constituents of hazardous waste burned in blast furnaces. These toxic metals would, like metals in the raw materials, be partitioned between the slag, pig iron, and the off-gases according to their physical and chemical properties. Those that leave the furnace in the off-gases will be pass through a high efficiency air pollution control system. It is conceivable that baseline emissions (i.e., without burning hazardous waste) of a metal could exceed eventually adopted limits. In this situation, the owners and operators may be

TABLE 4.18

COMPARISON OF HYDROGEN CHLORIDE
EMISSION DATA AND CONTROL EFFICIENCIES

Plant	HCl Input (lb/hr)	HCl Emissions (lb/hr)	Control Efficiencies (%)
A	62.3	<0.38	>99.4
B	15.5	<0.56	>96.4
C	1.3	<0.03	>97.7

required to burn hazardous waste with metal levels below specified limits instead of being required to meet mass emission rate limits.

As with toxic metals, there are no definitive data on the fate of chlorides in hazardous waste burned in a blast furnace. The chloring in chlorinated wastes would likely be converted to a metal halide which would probably be removed in the slag. Any HCl that is formed would have to pass through a bed of lime at some point near the top of the furnace. The burden contains limestone which is converted to lime as it travels downward in the furnace. This lime would likely neutralize most of the HCl formed from the combustion of chlorine-containing wastes. The air pollution control system would capture some of the HCl that might pass out of the blast furnace.

SYSTEM MODIFICATIONS NECESSARY TO BURN HAZARDOUS WASTES IN INDUSTRIAL FURNACES

In order to fire hazardous wastes in existing industrial furnaces certain modifications may be necessary, including:

- o Installing waste storage and handling facilities.
- o Installing equipment to characterize the wastes.
- o Installing equipment to pretreat the waste to improve its suitability for firing.
- o Adding a burner gun for firing the waste or replacing the existing gun with a multi-fuel burner designed to burn several fuels singly or in combination.
- o Upgrading combustion process controls to handle the waste fuel.
- o Providing waste feed rate, oxygen, and carbon monoxide monitoring to ensure that adequate destruction of the waste is achieved.
- o Updating the air pollution control system to meet imposed emission limits.

Waste Storage and Handling

A significant number of furnaces are used in industries which do not generate hazardous waste suitable for burning as fuel. Thus, waste burned in industrial furnaces is often received from off-site. Waste may arrive at the furnace location in tank trucks, rail tank cars, by barge, or other special means.

Although it is possible to feed the waste directly from these transport units, it is desirable to provide for some short-term on-site storage. Unloading facilities are needed to transfer the waste from the hauler to the storage tanks. Because the properties of these wastes are almost always very different from the raw materials and products handled at the furnace sites, existing unloading facilities are often inadequate for transferring hazardous wastes. Possible exceptions are facilities which use oil as the primary fuel for the furnace. It may be possible to use the oil unloading facilities at

some of these plants to transfer the waste to storage. As discussed earlier in this document, however, most cement, lime and lightweight aggregate kilns are fired by coal.

The requirements for hazardous waste storage and handling are being developed in a separate regulatory analysis by EPA OSW and should be consulted before designing and constructing such facilities. The subject will therefore not be discussed further in this document.

Waste Characterization and Pretreatment

As previously mentioned, essentially none of the potential hazardous wastes burned in industrial furnaces, particularly by the cement, lime, and lightweight aggregate industries are produced by these industries. The major exception to this generalization is in the pulp and paper industry where wastes from pulping processes are often burned as fuel in the kilns used for regenerating lime from lime sludge. Since the wastes are generated off-site, kiln operators can be selective as to which they will accept for burning. They are not forced to choose between firing the waste in the kiln or finding some alternative disposal. Specifications on the waste storage, handling, and firing problems; to prevent potential damage to process equipment; to protect product quality; and to meet environmental restrictions. Laboratory space, instruments, and personnel must be provided for characterizing the waste received from shippers. Samples of each shipment need to be analyzed to determine if wastes conform to the desired specifications. Waste not meeting the specifications would be rejected and returned to the shipper.

Although it is possible to eliminate the need for pretreatment by a judicious choice of specifications, to do so would severely limit the quantity of waste that could be fired in furnaces. An approach that is frequently used by furnace operators is to accept only those wastes that can be upgraded to fuel quality by one or more of the following simple, inexpensive pretreatment processes: waste blending, solids removal by inline straining, and thermal treatment. More expensive pretreatment is avoided by requiring that wastes meet established specifications or waste properties. The waste properties which affect furnace operation are those on which decisions regarding pretreatment are required. These properties include heating value, solids content, water content, ash, halogen, and sulfur content, and metals and toxic substances (e.g., PCBs) content.

A minimum heating value of 5000 to 8000 Btu/lb or greater is commonly used as a guide to define whether a waste stream is being incinerated or recycled for heat recover. When wastes are below this range, they can be blended with streams bearing higher values. Since the blending would be done batchwise, process equipment required would consist of an agitated tank and transfer pumps. Waste would be conveyed from a shipping vehicle into the agitated tank where it would be blended with waste of higher heating value. After the heating value of the blended waste is adjusted, it would be transferred to an appropriate storage tank or fed directly to the furnace.

Water in a liquid waste impacts a furnace in two ways. Free or undissolved water in a waste stream generally causes furnace burner pulsation and frequently leads to flame failure. Water also tends to lower the heating value of the organic fraction since a portion of the heat generated by its

combustion is consumed in vaporizing and heating the moisture to the furnace discharge temperature. Water removal is generally not practiced at furnace operations because the problems and costs associated with disposing of the generated contaminated water makes limiting the water content of waste more attractive. Most cement, lime, and aggregate plants do not have wastewater treatment facilities and those with treatment systems are limited in capacity (used mainly for treating scrubber water). A practical alternative to water removal is to develop specifications that would disallow any free water and limit dissolved water to less than 10% by volume. If a shipment of waste is received which contains water in excess of the specification, it may be rejected. Alternatively, it may be possible to blend it with another shipment or with waste stored on-site to meet the specification if the water content is not too high and the materials are compatible. Samples of the blended waste would be analyzed to ensure that the blended mixture met the water content specifications before it would be transferred to a storage tank or directly to the furnace burner.

High solids concentration in liquid wastes will increase the apparent viscosity; cause blockage of the burner nozzles; settle in waste solvent lines; and may (if not combustible and not incorporated in the product) increase particle loading to the air pollution control system. To minimize these problems strainers should be installed in the piping system and specifications should be adopted that require that the wastes be pumpable and that particle size be smaller than the strainer screen opening. In-line strainers consist of one or more mesh screen baskets housed in a vessel which may be one of a variety of geometric configurations. When the waste is passed through the strainer, the solid particles are trapped in the basket. Waste solvents often contain grit and debris such as o-rings, metal shavings, etc., that can damage pumps or plug burner nozzles. These can easily be removed by strainers. A 1/8-inch opening screen basket is well suited for removing such grit and debris. A 100 mesh screen is recommended for removing smaller particles which might plug a burner nozzle. Several strainer designs are commercially available differing mainly in the cleaning approach. A good choice for this application is a duplex strainer which permits the cleaning of one basket while another is on-line. It should be recognized that the strainer is not intended to remove large quantities of solids. Its purpose is to protect pumps and prevent plugging of small restrictions in the fuel train such as burner nozzles. As long as the waste is pumpable, however, solids removal should not be required. Pumpability of a waste containing a high solids content may be improved by blending the waste in the same manner as described above for increasing the heating value.

Halogens such as chlorine, bromine, iodine, etc., are commonly found in the raw materials used to produce cement and lime clinker and pig iron. They preferentially react with the alkali metals during production giving volatile halides at the higher furnace temperatures. As a result, very little halogens are found in most cement and lime clinkers and iron. The alkali halides are usually carried from these furnace in the gas stream. For this reason, additional chlorine is sometimes introduced with the cement kiln feed in order to reduce clinker alkali content. The potential benefits and problems associated with burning halogenated wastes in cement and lime kilns were discussed earlier in this document. The raw materials used in aggregate (lightweight and asphaltic concrete) production are not as alkaline and contain much less halogens than those for pig iron, cement, and lime (Reference

8) and the capacity of the aggregate product for forming alkali halides is much less than that for pig iron, cement, and lime clinker. As a result, the likelihood that chlorine will exit the furnace as HCl is much greater for aggregate (lightweight and asphaltic concrete) production. This can have a major impact on the air pollution control systems for these furnaces as most are susceptible to HCl corrosion.

The most practical means of protecting process equipment and product quality against the deleterious effects of high halogen content is to establish limits on the levels that will be accepted. Choice of a suitable limit will be, to some degree, dependent on the type of process used for making the product. If suitable equipment is available, waste may be blended to meet the specification. This would greatly increase the quantity of wastes that could be accepted for burning. A simple blending arrangement like that described above for increasing the heating value of waste may be used.

Viscosity is important to waste firing in two regards. A primary concern is that the waste be pumpable. Reasonable pump designs and piping pressure drops require limitations of waste fuel viscosity at about 10,000 standard saybolt units. Also, if the viscosity of the waste fired is too high, it will not be possible to atomize the liquid into droplets small enough to oxidize completely. Good atomization can usually be achieved if waste fuel at the burner is less than 750 standard saybolt units (Reference 20). This figure is only a generalization as some can handle more viscous fluids while others cannot handle liquid approaching this viscosity. If wastes must be pumpable to be accepted for burning, viscosity related problems can be satisfactorily handled by keeping the waste heated, by blending with other wastes, or a combination of the two. Viscosity can be reduced by heating the waste with tank coils or in-line heaters. However, 400° to 500°F is the normal limit for heating to reduce viscosity since pumping a hot tar or similar material becomes difficult above these temperatures. Prior to heating a liquid waste stream, a check should be made to ensure that undesirable reactions such as polymerization, nitration, oxidation, etc., will not occur. If preheating is not feasible based on these considerations, a miscible liquid of lower viscosity may be added to reduce the viscosity in a simple blending tank like that described above for heating value adjustment.

Where viscous materials are to be burned, provisions for both thermal treatment and blending to regulate the viscosity may be necessary. Storage tanks will be insulated with a two- or three-inch layer of suitable insulation and heated with a side-mounted steam or electric heat exchanger, or steam coils installed near the floor. A gear pump designed for expected temperature and viscosity is recommended to transfer the waste fuel. The pump and piping should be insulated and heat traced to prevent the waste from cooling in the lines.

Hazardous waste also often contains various levels of ash, sulfur, toxic substances such as PCB, and heavy metals such as lead, cadmium, mercury, etc. The atmospheric discharge of some of these materials is generally limited by local, state, or federal regulations. One regulatory approach is to restrict the levels of these materials in the wastes being burned. Pretreating these wastes by means other than blending to meet the imposed specifications is considered economically impractical.

Waste Burner

For many furnaces a burner gun must be installed to fire wastes. As previously mentioned in Section 3, pulverized coal is the primary fuel for the vast majority of cement, lime, and lightweight kilns. The burner for firing this fuel is essentially an 8- to 24-inch diameter tube which projects from 5 to 10 feet into the kiln. Pulverized coal is blown into the kiln with the primary combustion air through this tube. Although this type of burner may be suitable for firing some gaseous and solid wastes, it is extremely inappropriate for liquid wastes. Inasmuch as liquid wastes must be vaporized before burning can take place, they must be atomized to achieve efficient combustion. The pulverized coal burner has no atomizing mechanism. Liquid wastes could be sprayed onto the coal before it enters the pulverizer, but this practice would cause significant equipment and safety problems. Thus, an atomizing gun must be installed to fire liquid wastes in pulverized coal fueled kilns. Although gaseous waste can be simply piped into the coal burner, generally a center-feed gas gun is installed to give better mixing of the waste with the primary combustion air and to achieve good flame pattern control. This gun is essentially a 1- to 2-inch diameter tube that is tapered at the firing end. Slots are provided in the tapered end to distribute the gas in order to enhance mixing with the combustion air. It is typically located in the center of the coal tube. It may or may not run the full length of the coal burner tube. It is unlikely that gaseous wastes would be burned in a kiln, as those wastes are typically captive waste and are burned as generated.

A gun must also be provided to fire waste in furnace equipped only with natural gas burners. These burners cannot be used for firing liquid and solid wastes. Although it is possible to simultaneously fire a waste gas with natural gas through this type gun, a separate waste gun is generally used in order to provide better flame control. Thus, a separate waste gun is also needed for natural gas fueled kilns.

Furnaces equipped with oil guns may not require a separate gun to fire waste. A few furnaces employ a combination of burners: oil and gas; oil and coal; or oil, gas, and coal. These burners may fire the fuels separately or in combination. Typically, only one fuel (generally oil or gas) is fired at a time and the capability of firing the other fuel(s) is maintained for standby purposes if the supply of the primary fuel is temporarily unavailable. For furnaces with oil burners, some liquid wastes may be fired using the standby oil gun. The oil supply line is disconnected from the burner and plumbed to the waste handling system. In order to use it for this purpose, however, the waste viscosity, solids content, and particle size must meet the design specifications for the gun. Since the burner turndown is generally low (about 5:1) for this type of gun, its use at low feed rates may be limited due to poor atomization. Another approach for using an existing oil gun is to blend wastes and oil if they are compatible. This approach may be used even when the burner is not a combination burner. However, oil is often used for startup purposes and furnace temperatures are low at startup. DREs of FOHCs in an oil/waste mixture would likely decline, so oil/waste mixtures should not be used for startup. Where the waste cannot be fired in the oil gun, a separate waste gun must be provided.

Regardless of which gun is used, it is crucial that it be suitable for the waste being fired and that it be operated in such a manner as not to interfere significantly with the overall performance of the furnace burner. The performance of the burner is critical to the operation of a furnace, whether the product is cement, lime, lightweight aggregate, or any of a wide variety of other materials. Both the furnace productivity and furnace thermal efficiency in terms of heat input per unit product output are functions of the furnace burner. Large variations in both productivity and thermal efficiency in the same furnace under identical conditions can be caused by changes in the shape, length, and location of the flame in the furnace. Even product quality can be affected by these changes. Thus, the entire plant production capacity will be affected by the ability of the burner to provide an optimum flame shape, length, and location. It is, therefore, extremely important that the firing of hazardous waste not alter this optimum flame pattern. When a waste is fired through a gun originally installed in a furnace to burn conventional fuel, the waste properties (e.g., heating value, viscosity, volatility, solids content) should not differ from the conventional fuel enough to significantly alter the flame pattern. Similarly, a burner installed for firing a waste to supplement the heat input provided by a conventional fuel must be designed and located such that the flame pattern is not altered significantly. Also, the flame should not impinge on the furnace wall. Because of heat radiation from the combustion zone of the furnace, an air-cooled jacketed waste gun may be necessary to prevent the waste from pre-volatilizing inside the gun. When the waste/fuel ratio is large, it may be necessary to replace the furnace burner with a combination burner to achieve a good flame pattern.

For liquid waste, good atomization may be achieved by:

- o High pressure mechanical atomization
- o Low pressure air atomization
- o High pressure air atomization
- o High pressure steam atomization

In general, mechanical atomization of the waste at proper temperature and pressure through the use of a correctly sized nozzle tip will provide complete combustion and a good flame pattern. In the simplest form, the waste is fed directly to the nozzle but turndown is limited to 2.5:1 to 3:1 since the degree of atomization drops rapidly with decrease in pressure (Reference 20). In a modified form involving a return flow of liquid, turndown up to 10:1 can be achieved. Major disadvantages of mechanical atomization are:

- o Erosion of the nozzle orifice
- o Tendency to plug with solids or liquid pyrolysis products
- o Potential for spraying hazardous materials over a large area in the event of a rupture of plumbing to the gun due to the high operating pressures

Air or steam atomization are also suitable for firing liquid hazardous waste in a furnace. These types of atomization will likely require addition of air or steam supplies if these are not available. Packaged systems are available to supply these commodities. The choice of low pressure air, high pressure air, or steam atomization will depend to a large extent on the waste

properties and the amount of air that can be tolerated. Low pressure atomization can satisfactorily atomize low viscosity wastes but with a high air requirement. Atomization air requirements vary from 370 to 1000 ft³/gallon of waste liquid fired. This air must be balanced with the primary combustion air supplied with conventional fuel to maintain an optimum flame. For high viscosity wastes, high pressure air or steam atomization, which is more expensive, is generally required.

Because of the importance of controlling the flame pattern in a furnace, not all types of burner guns are suitable for co-firing hazardous waste with the conventional fuel. Equally important, some types are unacceptable because of the potential environmental risk associated with their use. The simple tube or pipe with no atomizing nozzle is an example of an unsuitable waste burner gun. For liquid wastes, which constitutes the major portion of hazardous wastes fired in furnaces, good atomization is essential for complete combustion of the waste and for good flame control. The rotary cup atomizing gun is also not suitable for firing waste in most industrial furnaces. These guns produce larger droplets than atomizing burners and their flame pattern control is poor when used over a wide range of flow.

Combustion Process Control and Safety Shutdown

In order to safely fire hazardous waste in an industrial furnace, it may be necessary to improve the fuel combustion controls. Since there is a variety of control instrumentation, the degree of equipment change is site-specific. For essentially all rotary kilns used in the lightweight aggregate industry, fuel combustion control is limited to manual adjustments of the fuel and combustion air feed rates by an operator. Similarly, for open hearth furnaces and many of the older kilns used in cement and lime production, fuel firing is controlled manually.

In order to maintain satisfactory manual control of the fuel combustion in the kiln, the operator must observe several parameters. The flame pattern and the color of the clinker are noted as these can indicate that the proper combustion is being obtained. Also, the temperatures of the gases and clinker are measured at various points. Gas samples are periodically withdrawn from the kiln exhaust for determination of oxygen and carbon monoxide levels. Draft pressure developed by the fans moving air through the kiln, preheaters, and/or precalciners are observed. Armed with these observations, the operator generally is able to manually control the fuel combustion to safely produce quality clinker.

Although hazardous waste may be safely fired with manual control of the fuel and combustion air feed rates, automatic termination of waste feed is deemed necessary to prevent the release of hazardous materials to the environment in the event of flameout, other combustion process upsets, or air pollution control device failure. This requires the installation of a flame scanner. This device senses ultraviolet radiation from the flame. When the flame is lost, the scanner signals an automatic valve in the waste feed line to close, immediately terminating the waste to the burner. An alarm should also be added to warn the operator of the waste fuel shut off. It is also deemed necessary to provide continuous monitoring of carbon monoxide levels in the furnace exhaust gases to ensure that proper air/fuel stoichiometry is maintained to achieve adequate combustion of the hazardous waste. An oxygen

analyzer is also necessary to correct the carbon monoxide readings for dilution by air inleakage into the flue gases. A carbon monoxide analyzer linked to an automatic waste field shut-off system could quickly terminate waste feed in the event that poor combustion conditions cannot be eliminated by equipment adjustments. Additional discussion of oxygen and carbon dioxide monitoring is contained in a later subsection.

Because of the ever increasing cost of energy, industrial furnaces constructed in the last five years are generally equipped with semi-automatic process controls which incorporate logic and instrumentation for managing fuel combustion. These systems vary considerably in capability with the more sophisticated ones including metering of the fuel(s) and automatic adjustment of fuel:air ratio using continuous oxygen and/or carbon monoxide monitoring. The controls may be set up to control single fuel or multiple fuel firing. They are usually based on known heating values or air:fuel ratio requirements for each fuel.

The extent to which the semi-automatic control systems must be modified to fire hazardous waste depends on the system design and desired control options. If a waste is to be co-fired with a conventional fuel and is to be restricted to less than about 25% of the total heat input, metering of waste fuel into the furnace by the process controls is not needed. The furnace could be manually baseloaded on the waste fuel with the process controls being used to manage process savings by adjusting the conventional fuel and combustion air flows. This would require no significant change in the combustion controls. Similarly, if the system is designed for multiple fuel firing, the required changes may be as minor as reprogramming the microprocessor to account for the heating value different between the waste to be fired and the fuel being replaced by the waste.

Significant changes to a furnace's combustion controls may be needed when independent modulation of the waste and conventional fuel feed rate is required. Two such waste/fuel burning configurations are identifiable. Independent modulation is needed to fire either 100% waste fuel or 100% conventional fuel when the properties of the waste are very different from those of the conventional fuel, e.g., liquid waste and pulverized coal. Similarly, co-firing at more than 25% of the total heat input with a conventional fuel would require considerable changes to the control instrumentation. Both these options necessitate additional flow control valves and may require a new microprocessor. An automatic flame supervision system is considered essential whenever one is not included as part of the combustion control instrumentation.

Waste Feed Rate Monitoring

It may be necessary to restrict the flow of some highly toxic waste streams to a small fraction of the total fuel input to ensure adequate destruction of a POHC. If such restrictions are adopted, waste feed rate monitoring will be needed. Additionally, a trial burn may be required to demonstrate the capability of a furnace to achieve an adopted DRE. The quantity of POHC being fed to the furnace is needed for the DRE determination. Where metering pumps or feeders are employed to convey the waste to the burner, a continuous measurement of the feed rate may be obtained by modifying these devices to provide a rate-dependent signal to some type of recording device.

The following paragraphs some of the more useful flow meters that may be used for this application. Detailed information on these and other flow meters can be found in References 21 through 24.

There is no universally applicable flow meter. The proper selection depends on the waste properties, acceptable pressure drop, required accuracy, and flow range to be measured. It is critical that the operation of the flow meter be compatible with the waste. Properties of chief concern in this regard include physical state, viscosity, solids content, electrical conductivity, and corrosivity. Pressure loss is an important consideration in that energy must be expended to overcome it. The flow meter must also be capable of measuring the flow to the specific accuracy over the flow range which it is used.

The orifice meter, the positive-displacement meter, the flow tube meter, the vortex shedding meter, and the turbine meter cover a wide range of applications for liquid and gaseous hazardous wastes. These instruments are moderately inexpensive, are capable of the needed accuracy, are of relative simple design, and can be used over a large range of flows. The orifice meter and flow tube are differential pressure type flow measurement devices. This type directly measures fluid velocity by measuring a differential head (pressure) across an obstruction in the flow stream which increases the velocity of the fluid, thereby decreasing its pressure. Flow equations relate the velocity change to the pressure change. In an orifice meter, the differential pressure between the upstream and downstream sides of an orifice plate is measured with pressure taps on either side of the orifice plate.

One disadvantage of the orifice meter for use in this application is that suspended matter in the fluid may build up at the inside of the orifice plate and affect its accuracy. This can be avoided by keeping the solids content low. If it is not practical to reduce the solids content, the flow tube may be used. The flow tube is basically a venturi without the downstream recovery cone. Because it does not restrict the flow to the extent an orifice plate does, it is applicable to streams with appreciable solids content. It has a very constant discharge coefficient and is considered to be highly reliable. It is not as expensive as the venturi but considerably more expensive than the orifice meter.

The positive-displacement flow meters have one or more moving parts positioned in the flow stream. The main devices are reciprocating piston, rotary piston, rotary-vane meter, and nutating disk. Of these, more nutating disk meters are probably used in than all the others combined. This device consists of a movable disk mounted on a concentric sphere. The disk is contained in a working chamber with spherical sidewalls and top and bottom surfaces that extend conically inward. The disk is restricted from rotating about its own axis by a radial partition that extends across the entire height of the working chamber. Each complete movement of the disk displaces a fixed volume of liquid. The liquid enters through an inlet port and fills the spaces above and below the disk, which fits closely and precisely in the measuring chamber. The advancing volume of liquid moves the piston in a nutating motion until the liquid discharges from the output port. The major limitation of this type of flow meter is its sensitivity to grit.

With a vortex shedding meter, the fluid stream is forced past an obstruction (shedding bar) which sets up vortices (eddies) in the fluid. These vortices cause vibrations in the shedding bar which are proportional to the flow. A piezoelectric crystal converts these vibrations to a voltage that is amplified and transmitted to an electronic scaling module. Advantages of these instruments include accuracy, no moving parts, and relatively low price.

The turbine meter is a mechanical-type measurement instrument. It operates on the turbine principle, i.e., the volume is measured by the movement of a wheel or turbine type of impeller. The blades of the turbine, which are positioned within a chamber, rotate as the fluid passes through them. The rotor can be positioned so that it can be driven by radial or axial flow or a combination of both. The rotor's motion can directly drive a register. This device can be used to measure continuous high gas flow rates with minimum pressure loss. It is not well suited for liquids containing appreciable quantities of solids.

Oxygen and Carbon Monoxide Monitoring

The test results presented earlier in this section demonstrated that furnaces used in cement, lime, and aggregate production are capable of a 99.99% DRE. However, good combustion conditions must be maintained to ensure high DREs. Like all combustion devices, industrial furnaces are subject to operating excursions. An acceptable DRE must be maintained during these excursions to prevent the emissions of hazardous materials to the environment or else the waste feed must be shut off. Continuous monitoring of DRE is impractical because of the complexity (and cost) of the sampling and analysis required. Consequently, some surrogate indicator of the combustion performance is needed. Combustion performance depends on operating parameters such as temperature, feed rate of waste, and air flow rate, but monitoring these operating parameters does not indicate what is actually being accomplished in the furnace in terms of the waste destruction.

Monitoring oxygen and/or carbon monoxide levels in the flue gases does give a continuous assessment of the effectiveness of combustion. (For furnaces intentionally operated under reducing conditions, such as a blast furnace, CO monitoring as a surrogate indicator of waste combustion performance is not practical. However, the off-gases from reducing furnaces are typically burned in other combustion devices where CO monitoring as a surrogate indicator of combustion performance is practical). Generally accepted combustion theory holds that low CO (carbon monoxide) flue gas levels are indicative of a furnace operating at maximum combustion efficiency. Operating at maximum combustion conditions ensures minimum emissions of unburned (or incompletely burned) organics. In the first stage of combustion of hazardous waste fuel, the POHCs are immediately thermally decomposed in the flame to form other, usually smaller, compounds termed PICs. In this first stage of combustion these PICs are also rapidly decomposed to form CO.

The second stage of combustion involves the oxidation of CO to CO₂ (carbon dioxide). The CO to CO₂ step is the slowest (rate controlling) step in the combustion process because CO is considered to be more thermally stable (difficult to oxidize) than other intermediate products of combustion of hazardous waste constituents. Since fuel is continuously being fired, both combustion stages are occurring simultaneously.

Using this view of waste combustion, the "destruction" of a POHC, and perhaps even the destruction of PICs, is independent of flue gas CO levels. Thus, CO flue gas levels cannot be correlated to DRE for POHCs and may not correlate well with PIC destruction. Low CO is an indicator of the status of the CO to CO₂ conversion process, the last, rate-limiting oxidation process. Since oxidation of CO to CO₂ occurs after destruction of the POHC and its (other) intermediate (PICs), the absence of CO is a useful indication of POHC and PIC destruction. The presence of high levels of CO in the flue gas is a useful indication of inefficient combustion, and at some level of elevated CO flue gas concentrations, an indication of failure of the PIC and POHC destruction process.

Instrumentation for both O₂ and CO monitoring of furnace flue gas is commercially available, is considered to be reliable, and is already installed on some industrial furnaces for combustion control. A variety of analyzers are used in these monitoring systems. These are reviewed in Reference 25 which also presents a list of vendors.

Although the analyzers could be used to manually shut-off waste feed, an automatic shut-off system consisting of a microprocessor-controlled shut-off valve is deemed more appropriate. The microprocessor could time-average the signals from the analyzers and signal the valves to stop the waste feed if a time-averaged CO limit is exceeded.

Air Pollution Control Equipment

All industrial furnaces are eventually are vented through some type of dust collection systems to control particle emissions. The efficiency of these systems was discussed in Section 3. Two important considerations in the RIA of firing hazardous wastes are:

- o How must these dust collection systems be modified so that hazardous waste can be fired without violating existing emission standards?
- o What further modifications may be required when RCRA regulation of hazardous waste firing in industrial furnaces is promulgated?

In order to answer the first question, an understanding is needed of the potential impact of firing hazardous waste on the furnace air emissions. This potential impact is directly related to the following waste constituents: ash, sulfur, trace metals, toxic substances (e.g., PCB), and halogens. These species exit the furnace as part of product, the waste solids and liquids, and/or in the exhaust gases. Any fuel conversion could increase the particle loading beyond the capacity of collection systems. This is not likely to occur with furnaces firing coal as waste fuels typically contain less ash than the 5 to 20% ash coal they replace. A waste with enough noncombustible material to create a particle emission increase for coal-fired furnace would create intolerable storage, pumping, and combustion problems.

As with conventional fuels, some of the sulfur in the hazardous waste fuel will exit the furnace in the exhaust gases as sulfur dioxide. The dust collection systems on most furnaces are not very effective for FGD (spent acid recovery furnaces are inherently equipped for SO₂ control). When burning conventional fuels, furnace operators find it more economical to use low

sulfur fuels than to install FGD. This being the case, there is little likelihood that high sulfur waste fuels will be accepted for incineration at all but spent recovery furnaces. Economics simply make it more practical to limit the waste sulfur content than to upgrade the existing dust collection system to provide for FGD.

Trace metal removal efficiencies of most furnaces dust collection systems have not been established. Because of the potential risk to the health of the population surrounding a furnace, a maximum allowable concentration of some of these metals in waste fuel could be established. This approach would eliminate the need to evaluate and possibly upgrade existing control systems to ensure that the public's health is adequately protected from exposure to atmospheric emissions of toxic metals.

Toxic substances like PCB, herbicides, and pesticides are generally limited to such low concentrations in the waste fuel by existing regulations that the need for further reduction by air pollution control equipment is impractical and unwarranted.

Firing halogenated wastes impact emissions in the following ways:

- o For cement and lime kilns and other furnaces which process alkaline raw material, it can change the resistivity of the dust and increase the exhaust dust loading if ESPs are used.
- o It substantially increases the HCl emissions from furnaces not processing alkaline raw materials which neutralize the HCl produced during combustion. These increased HCl emissions which can rapidly corrode the equipment used for controlling particle emissions from these furnaces.

The halogens released when halogenated wastes are broken into their constituent elements form volatile alkali halides in the cement and lime kilns and other furnaces processing alkaline materials. Some of the alkali halides, principally those formed from potassium (and possibly sodium), exit the furnace as gaseous fumes and later condense to dusts at the lower temperatures outside the furnace. This increases the dust loading to the control equipment. While the increased dust loading experienced by the dust collection system can result in increased particle emissions, it is quite feasible that the process and/or air pollution equipment operating conditions can be modified to adequately control emissions. Process changes, which can probably be employed, include reducing the insufflation rates of the recovered dust, reducing the rates at which leached dust is recycled, altering furnace drafts, and modifying furnace kinetics. The appropriate air pollution control device operation change needed to counter increased loading depends on the type of equipment being used. For an ESP, the increased loading may be offset by increasing the rapping frequency and modifying the voltage and current distributions. A baghouse is generally insensitive to slight changes in loading and increasing the cleaning frequency is one way of countering larger dust loading. For a scrubber, the increased loading can, in some applications, be offset by increasing the pressure drop or the amount of scrubbing water.

The raw materials used in making lightweight aggregate contain substantially less alkali metals than those used in making cement and lime. For this reason, the formation of alkali halides in lightweight aggregate kilns does not take place to the extent that it does in cement and lime kilns. As a result, corrosive acids, e.g., HCl, are generated when halogenated organics are fired in these kilns. Furthermore, the lightweight aggregate raw materials are not nearly as alkaline as those used in cement and lime production and is therefore, not as capable of neutralizing generated acids. Thus, these corrosive acids exit the kiln with the exhaust gases and enter the dust collection system which is generally a wet scrubber. This can create two potential problems. If the materials of construction are not resistant to acids, rapid corrosion will occur and the scrubber may have to be replaced. The second potential problem is that some states may have HCl emission limitations. If emission standards cannot be achieved with the existing scrubber, modification or replacement may be required.

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SECTION 5

COST ELEMENTS FOR THE REGULATORY IMPACT ANALYSIS

GENERAL

A key element of any regulatory impact analysis is an assessment of the potential economic impact of the various regulatory options being considered. This section identifies the major cost factors being considered by EPA in the assessment and provides selective cost data that are not being obtained by EPA from other sources. Regulations for the burning of hazardous wastes as hazardous waste derived fuels (HWDF) in industrial furnaces could potentially have an economic impact on waste producers, furnace operators, and regulatory agencies. This report is limited to those costs associated with burning the wastes in existing industrial furnaces. The actual economic impact analysis will be presented by EPA in a separate document.

The economic impact of burning hazardous waste in industrial furnaces is being analyzed by EPA in terms of how the various regulatory options alter the net fuel savings. This net savings is the difference between the credits associated with the fuel replacement and the increased capital and operating costs of firing waste over firing conventional fuels. Therefore, analysis of the economic impact of a regulatory option requires cost data for the significant elements of the credits and the incremental costs. The choice of cost elements will depend, to a large extent, on the level of detail specified of the analysis. These data are being obtained from a number of sources. The items being provided in this document fall into three major categories: (1) conventional fuel prices, (2) capital costs to modify the furnace system to burn a hazardous waste derived fuel (HWDF), and (3) the major operating and maintenance (O&M) costs associated with burning a HWDF.

The costs are presented so that they may easily be applied by EPA in its analysis as they are needed for each specific industrial furnace.

CONVENTIONAL FUEL PRICES

The prices of conventional fuels for industrial furnaces are to be used in the analysis of fuel replacement credits. Two sets of prices are considered. The first set is for 1982 which was the year covered by the OSW Burner Questionnaire survey of waste fuel users. The results of this survey will also be used in the economic impact analysis. The 1982 prices were:

- o Natural Gas - \$3.63 per million Btu
- o Distillate Oil - \$7.24 per million Btu
- o Residual Oil - \$4.62 per million Btu
- o Coal - \$1.09 per million Btu

The above 1982 fuel prices were obtained from documents published by the Energy Information Administration which is the branch of the Department of Energy responsible for collecting, compiling, and disseminating data on United States energy cost and usage. Natural gas and oil prices were developed from data provided in Reference 1; coal prices were based on data from Reference 2. These prices were representative of the national average values paid by industrial users in 1982.

One element of the RIA is to estimate the impact of the regulation on future HWDF burner operations. Therefore, a second fuel data set was needed; projected fuel prices for the period covered by the analysis. Fuel prices projected for the period 1985 through 2010, are listed in Tables 5.1 through 5.4. These projections were taken from Reference 3 currently being used by the EPA Office of Air Quality Planning and Standards (OAQPS) for its economical modeling as part of its effort to develop New Source Performance Standards (NSPS) for industrial boilers. Costs are updated periodically.

CAPITAL COSTS

When fossil fuel-fired furnaces are used to burn hazardous wastes, capital expenditures may be required for a number of system modifications, including: waste characterization and pretreatment, storage and handling facilities; fuel burners; combustion control instrumentation; waste feed metering; combustion gas oxygen and carbon monoxide monitoring; and air pollution controls for toxic metals and HCl emissions.

Cost data for waste storage and handling facilities have been generated by EPA for its economic analysis of treatment, storage and disposal facilities. Therefore, cost data for waste storage and handling are not presented here. Cost data for the remaining elements have been aggregated into three groups.

- o Waste characterization and pretreatment costs
- o Furnace modifications costs
- o Air pollution control device costs

The cost data presented below were developed from published data, vendor information, and engineering judgment. The following discussion includes the various elements comprising these three groups.

Waste Characterization and Pretreatment Costs

A combination of restrictions on waste characteristics and pretreatment of wastes is needed in order to avoid potential adverse impacts of firing hazardous waste on furnace equipment, product quality, and the environment. The potential problems of firing these wastes could be avoided merely by adhering to a stringent acceptance specification. Alternatively, one could accept wastes of unrestricted quality and depend on pretreatment to make them

TABLE 5.1

PROJECTIONS OF REGIONAL INDUSTRIAL NATURAL GAS PRICES^a
 (1982 dollars per million Btu)

Demand Region	1985	1990	1995	2000	2005	2010
1. New England	6.09	7.16	8.50	11.47	14.67	16.50
2. New York/New Jersey	4.90	5.41	6.42	8.67	11.09	12.47
3. Middle Atlantic	4.29	4.59	5.45	7.35	9.40	10.57
4. South Atlantic	4.82	5.70	6.76	9.13	11.67	13.13
5. Midwest	4.14	4.90	5.81	7.85	10.03	11.29
6. Southwest	4.27	4.63	5.50	7.42	9.49	10.67
7. Central	3.77	4.55	5.40	7.30	9.33	10.49
8. North Central	4.22	4.79	5.68	7.67	9.81	11.04
9. West	4.69	5.44	6.45	8.71	11.14	12.53
10. Northwest	4.90	5.08	6.03	8.15	10.41	11.72

^a Taken from Reference 3.

TABLE 5.2

PROJECTIONS OF REGIONAL INDUSTRIAL RESIDUAL FUEL OIL PRICES^{a,b}
 (1982 dollars per million Btu)

Demand Region	1985	1990	1995	2000	2005	2010
1. New England	4.09	5.15	7.32	8.92	11.11	12.78
2. New York/New Jersey	4.02	5.09	7.23	8.82	10.98	12.63
3. Middle Atlantic	4.01	5.09	7.22	8.81	10.97	12.62
4. South Atlantic	4.20	5.32	7.56	9.21	11.48	13.20
5. Midwest	4.34	5.44	7.73	9.42	11.73	13.50
6. Southwest	4.43	5.63	8.00	9.75	12.14	13.97
7. Central	4.40	5.51	7.83	9.54	11.89	13.67
8. North Central	4.25	5.40	7.67	9.35	11.65	13.40
9. West	4.49	5.71	8.10	9.88	12.31	14.16
10. Northwest	4.31	5.49	7.79	9.50	11.84	13.62

a 1.6% sulfur.

0.3% sulfur = 1.6% sulfur + \$0.68/MMBtu.

0.8% sulfur = 1.6% sulfur + \$0.35/MMBtu.

3.0% sulfur = 1.6% sulfur - \$0.45/MMBtu.

b Taken from Reference 3.

TABLE 5.3

PROJECTIONS OF REGIONAL INDUSTRIAL DISTILLATE FUEL OIL PRICES^a
(1982 dollars per million Btu)

Demand Region	1985	1990	1995	2000	2005	2010
1. New England	6.33	7.28	10.02	12.07	14.84	16.96
2. New York/New Jersey	6.27	7.22	9.93	11.97	14.71	16.82
3. Middle Atlantic	6.24	7.18	9.88	11.90	14.63	16.72
4. South Atlantic	6.08	7.03	9.68	11.66	14.34	16.39
5. Midwest	6.20	7.13	9.82	11.83	14.54	16.62
6. Southwest	6.08	7.06	9.72	11.71	14.40	16.45
7. Central	6.15	7.08	9.74	11.73	14.42	16.49
8. North Central	5.98	6.94	9.55	11.50	14.14	16.16
9. West	6.04	7.01	9.65	11.63	14.29	16.34
10. Northwest	6.04	7.01	9.65	11.63	14.29	16.34

^a Taken from Reference 3.

TABLE 5.4

PROJECTIONS OF REGIONAL DELIVERED INDUSTRIAL COAL PRICES^{a,b}
(January 1983 dollars per million Btu)

Demand Region	Coal Type	Sulfur Content (lb SO ₂ /MMBtu)	1985	1990	1995	2000
1. New England	Bituminous	<0.80	3.53	3.77	3.93	4.01
		0.80 - 1.08	3.42	3.67	3.80	3.92
		1.08 - 1.67	3.30	3.67	3.88	4.03
		1.67 - 2.50	3.25	3.68	3.85	3.96
		2.50 - 3.33	3.19	3.50	3.64	3.68
		3.33 - 5.00	2.68	3.98	3.29	3.48
		>5.00	2.94	3.21	3.41	3.56
2. New York/New Jersey	Bituminous	<0.80	3.34	3.52	3.62	3.74
		0.80 - 1.08	3.20	3.41	3.51	3.63
		1.08 - 1.67	3.10	3.42	3.54	3.66
		1.67 - 2.50	2.94	3.22	3.35	3.47
		2.50 - 3.33	2.87	3.14	3.22	3.31
		3.33 - 5.00	2.39	2.71	2.90	3.14
		>5.00	2.60	2.83	2.97	3.13
3. Middle Atlantic	Bituminous	<0.80	2.93	3.20	3.34	3.45
		0.80 - 1.08	2.77	3.05	3.17	3.30
		1.08 - 1.67	2.60	2.96	3.11	3.25
		1.67 - 2.50	2.40	2.74	2.88	3.00
		2.50 - 3.33	2.41	2.70	2.82	3.04
		3.33 - 5.00	1.98	2.50	2.81	2.96
		>5.00	1.77	2.14	2.32	2.54
4. South Atlantic	Bituminous	<0.80	2.92	3.32	3.47	3.66
		0.80 - 1.08	2.74	3.12	3.26	3.42
		1.08 - 1.67	2.55	2.60	3.03	3.31
		1.67 - 2.50	2.30	2.80	3.06	3.17
		2.50 - 3.33	2.64	2.53	2.55	2.70
		3.33 - 5.00	2.09	2.69	2.81	2.96
		>5.00	2.52	2.64	2.71	2.87
5. Midwest	Bituminous	<0.80	3.13	3.39	3.53	3.66
		0.80 - 1.08	2.94	3.22	3.37	3.48
		1.08 - 1.67	3.00	3.14	3.30	3.45
		1.67 - 2.50	2.70	2.97	3.08	3.21
		2.50 - 3.33	2.59	2.91	3.02	3.14
		3.33 - 5.00	2.18	2.46	2.78	2.91
		>5.00	2.23	2.42	2.52	2.67
	Sub-	<0.80	2.63	2.84	2.84	2.92
	Bituminous	0.80 - 1.08	2.63	2.84	2.84	2.92

TABLE 5.4--Continued

Demand Region	Coal Type	Sulfur Content (lb SO ₂ /MMBtu)	1985	1990	1995	2000
6. Southwest	Bituminous	<0.80	2.94	3.31	3.58	3.77
		0.80 - 1.08	2.89	3.26	3.51	3.69
		1.08 - 1.67	1.03	2.65	3.02	3.16
		1.67 - 2.50	2.71	3.09	3.19	3.21
		2.50 - 3.33	2.57	2.79	2.94	3.08
		3.33 - 5.00	2.95	3.17	3.39	3.46
		>5.00	2.82	2.84	2.97	3.07
7. Central	Bituminous	<0.80	2.77	2.97	3.09	3.14
		0.80 - 1.08	2.76	2.95	3.07	3.09
		1.08 - 1.67	2.95	3.23	3.34	3.46
		1.67 - 2.50	2.95	3.22	3.18	3.22
		2.50 - 3.33	2.94	3.15	3.27	3.40
		3.33 - 5.00	2.69	2.61	2.58	2.66
		>5.00	2.34	2.47	2.52	2.58
	Sub-Bituminous	<0.80	2.53	2.60	2.72	2.81
		0.80 - 1.08	2.53	2.60	2.72	2.81
		1.08 - 1.67	2.37	2.44	2.52	2.81
		1.67 - 2.50	2.43	2.48	2.63	2.54
	Sub-Bituminous	<0.80	1.52	1.62	1.59	1.75
		0.80 - 1.08	1.36	1.48	1.47	1.59
		1.08 - 1.67	0.86	0.92	1.01	1.12
		1.67 - 2.50	0.80	0.90	1.04	1.06
8. North Central	Bituminous	<0.80	1.64	1.87	1.92	2.01
		0.80 - 1.08	1.47	1.66	1.74	1.76
		1.08 - 1.67	1.29	1.34	1.35	1.38
		1.67 - 2.50	1.91	1.40	1.50	1.54
	Sub-Bituminous	<0.80	1.52	1.62	1.59	1.75
		0.80 - 1.08	1.36	1.48	1.47	1.59
		1.08 - 1.67	0.86	0.92	1.01	1.12
		1.67 - 2.50	0.80	0.90	1.04	1.06
9. West	Bituminous	<0.80	2.69	2.87	2.91	3.15
		0.80 - 1.08	2.63	2.76	2.82	2.98
		1.08 - 1.67	2.26	2.49	2.85	2.91
		1.67 - 2.50	2.53	2.78	2.60	2.65
	Sub-Bituminous	1.67 - 2.50	2.34	2.42	2.49	2.59
10. Northwest	Bituminous	<0.80	3.18	3.37	3.42	3.58
		0.80 - 1.08	3.10	3.24	3.30	3.42
		1.08 - 1.67	2.17	2.27	3.29	3.42
		1.67 - 2.50	2.80	2.81	2.78	2.80
	Sub-Bituminous	<0.80	2.06	2.14	2.29	2.36
		0.80 - 1.08	2.06	2.14	2.29	2.36
		1.08 - 1.67	2.05	2.10	2.29	2.36
		1.67 - 2.50	2.05	2.05	2.05	2.10

^a By ICF Inc., 1850 K Street, N.W., Washington, D.C., 20006.

^b Taken from Reference 3.

suitable for furnace firing. Depending solely on waste specifications would severely limit the quantity of acceptable waste. Conversely, depending solely on pretreatment is not technically and/or economically practical for many wastes. Therefore, the more practical approach for preventing waste-fuel burning related problems in industrial furnaces is a combination of waste restriction and inexpensive pretreatment processes (blending, straining of grit and debris, and thermal treatment).

Operational and toxicity related problems can be prevented by establishing limits on waste characteristics that significantly influence furnace equipment deterioration, product quality, and/or the environmental risk related to firing wastes. These characteristics include heating value, water content, solids content, halogen content, viscosity, ash, toxic substances such as PCB and metals. Limits on these characteristics could be used as criteria for accepting or rejecting waste from shippers or other plant operations. Each waste shipment would be analyzed to determine its acceptability as a fuel. Wastes and/or HWDF which could meet the specifications as received or after blending, in-line straining, and/or thermal treatment would be accepted for burning while those out-of-specification wastes would be rejected. Blending with either a conventional fuel or another compatible waste or HWDF would be used when possible to upgrade it to an acceptable quality, i.e., to reduce the concentration of metals, ash, sulfur, solids, water, or halogens, and/or to increase the heating value of wastes. In-line strainers would be used to remove grit and other large particles to protect pumps and to prevent plugging of the burner gun. Viscous wastes would be heated to make them pumpable and to permit proper atomization necessary for high combustion efficiency.

The cost to characterize and/or pretreat a waste to make it suitable for furnace firing offsets the money saved in fuel replacement. This section presents these offsetting costs which are required for an economic analysis of waste firing in industrial furnaces.

Waste Characterization

The use of waste specifications as criteria for eliminating unsuitable wastes requires characterization of the physical and chemical properties. Therefore, provisions must be made for the necessary analyses. Three approaches for obtaining the analyses are possible. One approach would be to perform all analyses on-site, which would require the acquisition of analytical equipment. Laboratory space for the needed equipment could generally be made available in an existing laboratory or by converting office space. Another option would entail using a commercial laboratory for the required analyses. A third approach, is to perform some of the simple analyses in the plant laboratory while contracting a commercial laboratory for the more sophisticated analyses. The third approach was used to prepare analytical costs. It was assumed that analyses for metals and toxic substances would be made by a commercial laboratory while the other analyses would be conducted in a plant laboratory. Sufficient plant laboratory space to conduct the simple analyses was estimated to cost approximately \$50,000. Equipment for determining compatibility, heating value, halogen content, sulfur, ash, water content, and specific gravity could be purchased for approximately \$25,000. Costs for having the analyses made by a commercial laboratory are estimated later in this section along with other waste firing related O&M costs.

Blending

Blending a waste with a conventional fuel or another higher Btu waste provides an inexpensive means of reducing the concentration of metals, ash, sulfur, solids, water, or halogens, while increasing its heating value for safe disposal in an industrial furnace. Blending wastes is generally a batch process. Grab samples taken from each waste lot are used to confirm that they can be upgraded to the established specifications by blending. The selected waste batch is transferred to the blend tank where it is mixed with another waste or fuel oil. Samples of the materials to be blended are tested for compatibility before mixing. After the materials are mixed, the blended waste is either pumped into a larger tank for storage or directly to the furnace burner(s).

The equipment should be designed to process wastes received from off-site as well as from other plant operations. A simple system would consist of an agitation tank plus pumps and piping. At minimum, it would use a 20,000 gallon tank which is large enough to hold a tank truck load (5,000-10,000 gallons) plus the higher quality blend stock. A typical blending operation might operate one shift per day for five days per week, blending 20,000 gallons per shift. This schedule is especially suited to plants that only run one shift per day such as asphalt concrete plants. The blending schedule includes time for sampling and analysis, transferring it to the blend tank, adding high quality material, mixing the tank contents, and pumping to longer-term storage.

The tanks should be equipped with nitrogen blanketing as it is sometimes required under individual State Implementation Plans to minimize evaporative losses. Two 100 gallons/minute gear pumps are required per blending tank - one for transferring materials into the tank and one for pumping them out. Gear pumps were specified because they afford a reasonably tight shut-off and prevent leaking when not in operation. A 100 gallons/minute capacity would be needed to achieve rapid transfer of materials (e.g., empty the tank in approximately 200 minutes). Two hundred feet of three-inch piping was assumed to be needed for connecting the blend tank to other storage and handling facilities.

The total installed cost of a system with only one 20,000 gallon blending tank was estimated at \$74,000 including pumps and piping. This system would provide up to 20,000 gallons of blended material per day or 100,000 gallons per week for the typical five day per week blending operation. It would produce 595 gallons per hour of waste fuel for continuous furnace firing. It was considered the minimum practical size for blending waste for furnace firing. Larger blending capacities may be obtained by using multiple units. Costs associated with blending larger quantities may be estimated by multiplying the price of one unit by the number of units required.

Straining of Solids

Liquid wastes with high solids content may cause several problems including increased apparent viscosity, burner gun nozzle blockage, obstruction of piping, and possibly increased particle emissions. Because of these potential problems, the solids content should be sufficiently restricted to provide a liquid that is pumpable and free of grit and debris. In-line strainers have

been used to protect pumps and burner guns in the event that large particles turn up in the waste. Blending may be used to reduce the solids level but other pretreatment techniques such as filtering are not generally practiced because of the expense and the need to dispose of the removed solids which would be classified as hazardous waste.

Typically, two different sets of duplex strainers are needed. A 1/8-inch opening screen basket is required to protect the pump used to unload the waste from truck and rail cars. A 100-mesh screen is common for removing smaller particles which might plug a burner nozzle. Dual or duplex basket strainers allow strainer cleaning without flow interruption and are relatively inexpensive. A strainer capable of handling 250 gallons/minute is recommended for the waste unloading line while the one protecting the burner gun may be 100 gallons/minute or less depending on the rate of fuel flow. Installed costs of these strainers are \$1,000 and \$6,000 for the 100 and 250 gallon/minute capacities, respectively. These 1982 costs are for all carbon steel construction.

Thermal Treatment

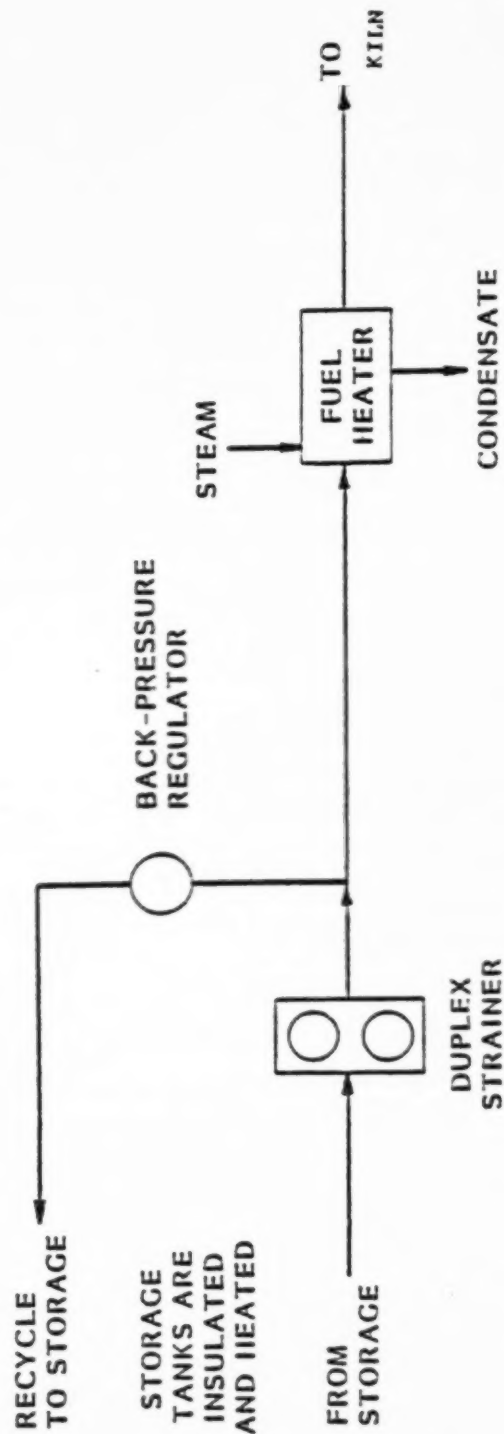
The techniques for introducing wastes into the furnaces are similar to those used with conventional fuel. The liquid must be sufficiently atomized to permit fairly rapid combustion. Although good atomization may be achieved with fluid viscosities as high as 5,000 standard saybolt units (Reference 4), it is common practice with standard fuel oils to limit viscosity to 750 standard saybolt units at the burner nozzle. Pumping problems also increase with increased liquid viscosity. The common practice with fuel oils is to maintain viscosity below 10,000 standard saybolt units for storage and transfer.

Blending is the preferred treatment process for adjusting the viscosity of wastes, but is not universally applicable due to incompatibility problems. When it is not possible to use blending, the viscosity may be reduced to acceptable levels by heating the waste. For purposes of developing thermal pretreatment costs, liquid wastes were categorized into three groups according to their viscosity:

- o Viscosity at ambient temperature <750 standard saybolt units.
- o Viscosity at ambient temperature is between 750 and 10,000 standard saybolt units.
- o Viscosity at ambient temperature >10,000 standard saybolt units, but <10,000 standard saybolt units, at 200°F.

If the viscosity was less than or equal to 750 standard saybolt units, the waste was assumed to be both pumpable and capable of being properly atomized. No thermal treatment would be needed for wastes in this category. Wastes of viscosity between 750 and 10,000 standard saybolt units are pumpable but must be heated to achieve satisfactory atomization. A fuel heater (shell and tube heat exchanger) may be used to raise the liquid temperature sufficiently to lower the viscosity to below the limit for good atomization (750 standard saybolt units). An equipment arrangement for this purpose is shown schematically in Figure 5.1. In this scheme, the waste is heated (by steam) as it passes through the heat exchanger before entering the waste burner. The only thermal treatment equipment items needed are the heater

FIGURE 5.1
EQUIPMENT DIAGRAM FOR THERMAL TREATMENT OF HAZARDOUS WASTES



and piping. The installed costs of this equipment are plotted as a function of waste feed rate in Figure 5.2 as Curve 1.

Wastes of greater than 10,000 standard saybolt units viscosity at ambient temperature must not only be treated to insure good atomization, but must also be kept hot enough to prevent them from setting up in the storage tank and piping. All waste received at the waste derived fuel storage facility must be pumpable. The equipment required to handle waste in this category include the heat exchanger for heating the waste as it is pumped to the furnace, provisions to recirculate the waste back to the storage tank, as well as items to keep the liquid warm:

- o Storage tank insulation
- o Storage tank heaters
- o Pipe tracing and insulation
- o Addition of a spare circulating pump

Installed costs of this equipment, excluding the storage tank insulation and heaters, are plotted as a function of waste throughput in Figure 5.2 as curve 2. Although waste storage tank costs are not part of the authorized scope of this study, the tank heater and insulation costs are provided. These costs are presented in Table 5.5. The storage tank heater for which costs are presented is the immersion steam coil type. Insulation costs are for three inches of fiberglass insulation.

Furnace Modification Costs

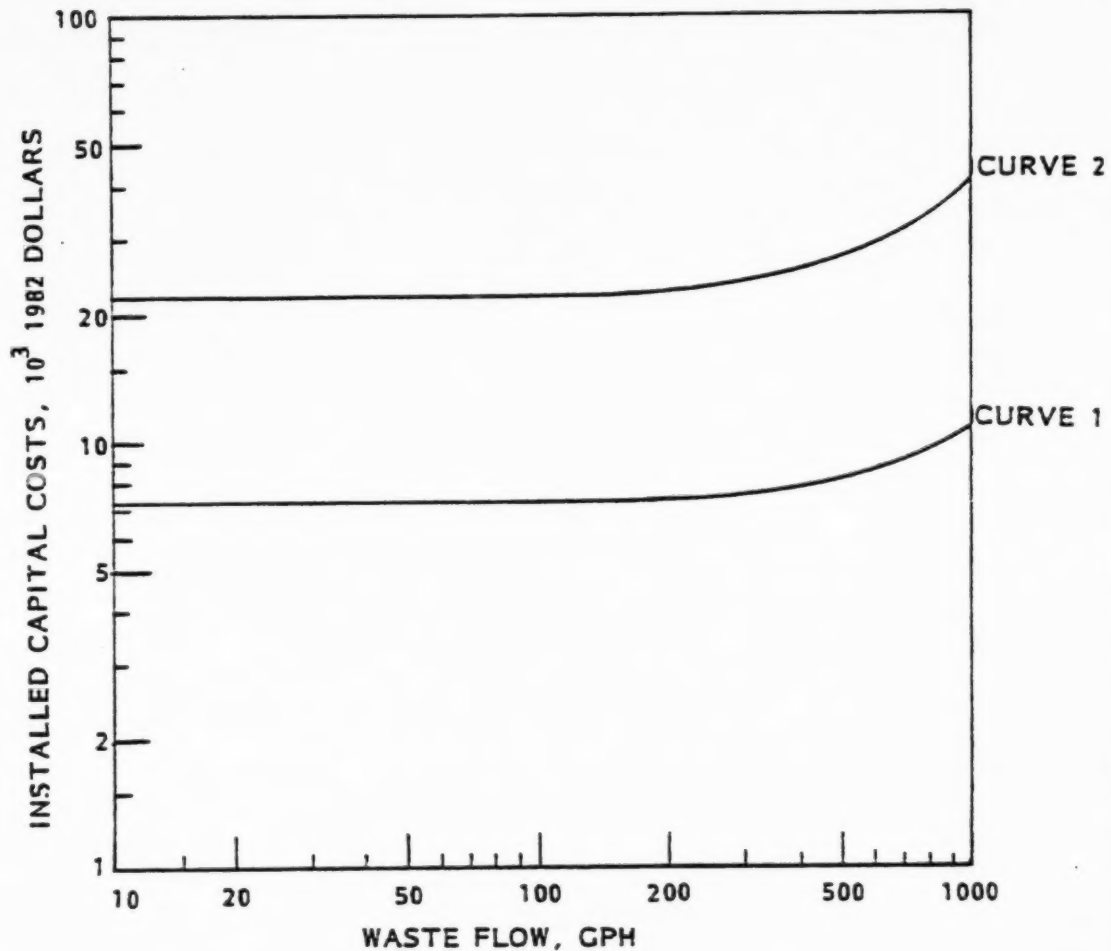
Burner Modification Costs

The cost for modifying an existing burner system to fire hazardous waste is site specific. It depends on the existing burner type and capacity, type of conventional fuel fired, properties of the waste, and quantity of waste to be fired. The least expensive approach is likely to be taken. Some furnaces require only that a burner gun be replumbed to fire the waste. This would not require significant capital expenditure. In other instances, the hazardous waste could be blended with the conventional fuel and fired with no modifications to the burner. This approach could be used when the waste and conventional fuel are compatible or when burning liquid waste in the fuel oil-fired burners.

Costs given below are for providing the necessary burner components to fire a liquid hazardous waste in natural gas, oil and combination fossil fuel-fired furnaces. It was assumed that the waste was piped to the burner, i.e., costs do not include a fuel handling train. The figures were obtained from burner vendors and, therefore, may be different from actual costs because some furnace operators may fabricate their own waste burners.

One approach to firing waste materials is to install a dedicated waste burner gun in or adjacent to the conventional fuel burner. Waste is simply atomized into the primary fuel (natural gas or oil) flame envelope. This arrangement may be used for co-firing waste up to 30% of the total heat input. A good quality atomizing burner gun, properly sized for the range of waste flows can be purchased for \$10,000 to \$15,000, depending on the length of the burner. The total installed cost in 1982 dollars, including plumbing and flame safeguard system, is from \$20,000 to \$30,000.

FIGURE 5.2

INSTALLED CAPITAL COSTS FOR THERMAL
TREATMENT OF HAZARDOUS WASTES

CURVE-1- PUMPABLE AT AMBIENT TEMPERATURES BUT MUST BE HEATED FOR GOOD ATOMIZATION.

CURVE-2- MUST BE HEATED TO KEEP PUMPABLE AND TO OBTAIN GOOD ATOMIZATION.

TABLE 5.5

STORAGE TANK HEATER AND INSULATION INSTALLED COSTS

Tank Size (gallons)	Heater Costs (1982 dollars)	Insulation Costs (1982 dollars)
500	11,000	430
1,000	12,000	660
2,000	14,000	960
5,000	17,000	1,310
10,000	19,000	2,850
20,000	38,000	3,640

It is assumed that the process controls for modulating an additional burner are not necessary. The furnace could be manually based loaded on the waste fuel with the existing process controls being used to modulate the conventional fuel and combustion air flows. Based on the detailed study of the furnaces and their current practices, this assumption is valid and should extend to other furnace types.

Carbon Monoxide and Oxygen Monitoring

Continuous monitoring of carbon monoxide levels in the exhaust gases of a furnace burning a hazardous waste is deemed necessary to ensure that good combustion conditions are maintained to provide adequate destruction of the POHCs and PICs. Oxygen monitoring is required in conjunction with CO monitoring to adjust CO levels to a common excess air rate indicated by excess oxygen content in the flue gas. Correcting CO levels to a common flue gas oxygen content avoids the problem of having (otherwise) high CO levels diluted by large quantities of excess air. Some furnaces are already equipped with oxygen and/or carbon monoxide monitoring to save fuel costs through increased combustion efficiency, and/or for safety purposes. Most of the CO monitors installed on cement kilns are for detecting explosive or unburned hydrocarbon concentrations and are not designed to measure the low CO levels typical of optimum combustion conditions (0-5000 ppm). These devices typical operate in a concentration range of 0-10%.

The cost of CO and oxygen continuous emissions monitoring systems for a "clean" environment is considerably less than that for a "dirty" application. There is not a clear distinction between a "clean" and a "dirty" exhaust gas. For purposes of this analysis, however, a "dirty" exhaust gas application is defined as one that requires an extractive monitoring system in which the extracted sample must be treated to reduce the gas temperature, particulate loading, and moisture content. A sample location in the backend of a dry cement kiln is an example of a "dirty" application. A "clean" application example is a sample location downstream of the air pollution control device.

Typical costs for oxygen and carbon monoxide continuous emission monitoring systems for clean applications are \$15,000 and \$20,000 respectively (1982 dollars). These costs include analyzers, sample transport system (if extractive type unit), chart recorder and installation. If an automatic data reduction system is desired an additional cost of \$20,000 would be incurred (Reference 5).

For a more hostile environment, the cost of the continuous emission monitor systems (CEMS) will be significantly higher due to the added cost of specialized sample conditioning equipment. Providing an unadulterated sample that is compatible with the analyzers is more difficult in sample locations where there are extremely high temperature, moisture, and particulate loading. The cost of special equipment needed to adjust these stack gas characteristics increases with the number of characteristics that must be adjusted and the degree of adjustment required. For purposes of this analyses it will be assumed that the conditioning system must treat a flue gas that requires a water-coded probe and has a high moisture content and particulate loading.

Waste Feed Metering

If limitations on the hazardous waste feed rate are established based on the trial burn results, capital outlays for flow metering will be necessary. Two types of liquid flow measurement devices well-suited for this application are the positive displacement meter and the orifice meter. These units are relatively inexpensive and are applicable to a wide range of wastes. Either device can be obtained for approximately \$2,000 to meter flows up to 7,000 gallons per hour, which is the maximum flow rate one would anticipate for waste fed to the industrial furnaces being considered. Some type of recording device is needed to provide a permanent record of the waste feed rate. A simple \$2,000 chart recorder would serve this purpose adequately. Installation is likely to cost an additional \$1,000. Thus, the total installed cost of a liquid waste feed metering system will be approximately \$5,000 in 1982 dollars.

Air Pollution Control Cost

The issue of particulate matter, metals, and hydrochloric acid emissions from a furnace burning hazardous waste need to be considered on a source specific basis. The furnaces considered in detail in this study are already equipped with air pollution control systems for removing particulate matter from the furnace exhaust gases. It is possible that most of these systems will be able to meet adopted toxic metal emission standards should they be adopted. It is assumed that metals will be controlled by either particulate matter controls or waste fuel specifications. Furthermore, hydrochloric acid emissions may not be a problem if operators do not burn chlorinated waste at chloride levels that could cause the adopted emission levels to be exceeded. However, should a furnace operator chose to burn HWDF's that would cause acceptable health risks from toxic metals and hydrochloric acid emission to be exceeded, additional air pollution controls would be required.

The type of control system that would be installed will vary depending on the emission limitations eventually adopted and the exhaust gas parameters. For study purposes, costs are presented for a combination venturi scrubber for particulate matter removal followed by a packed bed absorber for HCl removal. The venturi scrubber will remove metals that are contained in the exhaust gases as particulate matter while those in the vapor state will be removed by the packed tower. Costs are presented as a function of furnace exhaust gas flow. Factors are also given to estimate costs if only metals as particulate matter or HCl removal is required. Other combinations of toxic metals and HCl removal are available, but the venturi/absorber is a practical technique that adequately represents the costs element needed to assess the economic impact of pollution control equipment that may be required. This approach is somewhat conservative in that the packed tower may not be necessary in all cases to accomplish similar results.

Installed cost for a combination venturi/absorber system may be estimated from the following equation:

$$\text{Cost} = 96 \times Q^{0.8164}$$

where: Q = the exhaust gas flow rate in acfm

This system includes a quench tower to lower the exhaust temperature from 550°F to saturation, a venturi scrubber for particle collection, acid gas absorber, caustic recycle system for neutralizing the scrubber water, ID fan, stack, and auxiliaries. The assumed materials of construction are:

- o High-nickel-alloy quencher and venturi throat
- o High-grade, chemically resistant, high-temperature fiberglass shell for the cyclonic separator and packed tower
- o Polypropylene tower packing
- o Inconel or Hastelloy fan wheel with rubber-lined steel housing

If no venturi scrubber is needed for metals control, the total system cost is reduced by 15%.

If acid gas absorption is not required, the total system cost is reduced by approximately 40%.

Costs represented by the above equation are also based on purchased equipment cost contained in Reference 6 with installation assumed to be 100% of the purchased cost. These costs are typical for a venturi pressure drop requirement of 30-inches water column, which should be adequate for this application. They are indexed to 1982, the year covered by the OSW Burner Questionnaire which is being used in the RIA.

OPERATING AND MAINTENANCE COSTS

The operating and maintenance costs for a furnace operation are likely to be greater when 100% waste fuel or a combination of waste and conventional fuels are burned than when 100% conventional fuels are burned. Additional storage and handling facilities are typically required to fire wastes and these must be maintained. If agitators or nitrogen blanketing are used, operating costs will increase. When the wastes require pretreatment before burning, additional O&M costs are incurred. There may also be increased furnace O&M costs as a result of adding a waste burner gun. However, these costs are not expected to be significant.

The equipment needed to meet the regulatory requirements that are eventually adopted will likely further increase the O&M costs associated with firing hazardous waste over those associated with firing conventional fuels. Air pollution control devices added to meet emission limitations will increase the O&M cost burden. Further costs will be incurred to maintain oxygen and/or carbon monoxide monitoring and waste metering. O&M costs associated with waste metering are, however, expected to be insignificant.

This section presents O&M costs for the following of the aforementioned items:

- o Waste Characterization and Pretreatment
- o Furnace Modification Costs
- o Air Pollution Control Devices

Increased O&M costs, associated with waste storage and handling other than pretreatment, are provided by others.

Waste Characterization Pretreatment O&M Costs

The annualized O&M costs associated with waste characterization and the three pretreatment processes (blending, straining, and thermal treatment) for which capital costs were given earlier in this section are presented here.

Waste Characterization Cost

As previously discussed, furnace operators can avoid high O&M costs to upgrade hazardous waste by limiting the materials that they will accept to those that require only blending, straining, or thermal treatment to make them suitable for fuel. This practice requires that each shipment of waste be analyzed to determine if it meets established specifications. Wastes should be analyzed for heating value, halogens, sulfur, water, ash, specific gravity, toxic metals, and PCBs. These analyses may be made by a plant chemist or by an independent commercial laboratory. The need for rapid results requires the analyses to be made at a laboratory near the plant. Since metal and PCB determinations require more sophisticated analytical procedures and equipment, it is assumed that these are performed at a nearby commercial laboratory while the other analyses are completed in the plant laboratory.

O&M costs for the waste characterization approach described above will, therefore, include simple charges for the analyses performed by the outside firm and the labor and other costs associated with the plant's waste analyses. Assuming 30 samples are analyzed per week for PCB by gas chromatography and up to five metals by atomic absorption, the annual charges by the outside laboratory is estimated at \$156,000 in 1982 dollars.

Labor for waste characterizations by the plant's chemist and his supervisor is estimated at \$26,500. Maintenance and indirect O&M costs are taken at 22.2% of the capital costs of the laboratory space and equipment. These costs are \$16,600 which brings the total annual waste characterization O&M costs to \$199,100 in 1982 dollars.

Pretreatment

Certain elements of the pretreatment costs are generic to all three processes. These are listed in Table 5.6. As shown in Table 5.6, the O&M costs include both direct and indirect components. The direct components include operating labor, maintenance, and utilities. Operating labor is taken at \$9.75/manhour. Estimated labor requirement is two manhours for waste blending. For straining and thermal treatment, it is assumed that no operating labor is needed. Supervision is estimated at 15% of the total operating labor costs (Reference 7). Maintenance requirements are difficult to predict accurately for these types of operations. For such situations, maintenance is generally taken at 2-6% of the capital costs (Reference 8). Five percent was used in these estimates.

Utilities required for the pretreatment processes include electricity to drive pumps and the blend tank agitator, nitrogen for blanketing the blend tanks, and steam for the thermal pretreatment. Pump power consumption is estimated from the following equation:

TABLE 5.6

COMPONENTS OF ANNUALIZED O&M COSTS

Direct Operating Costs	Cost Factor ^a
Operating Labor	
- Operator	\$9.75/manhour
- Supervisor	15% of operator
Maintenance	5% of capital costs
Utilities	
- Electricity	\$0.05/kilowatt-hour
- Steam	\$6.00/1,000 pounds
Indirect Operating Costs	Cost Factor ^a
Overhead	80% of operating labor and main- tenance labor
Property Tax	1% of capital costs
Insurance	1% of capital costs
Administration	2% of capital costs
Capital Recovery Costs	0.132 (using $i =$ 10% and an equipment life of 15 years)

^a All costs are in 1982 dollars.

$$\text{kilowatt-hours} = \frac{0.746 (\text{gpm}) (\text{hd}) (\text{sg}) H}{3960 n}$$

where: gpm = flow rate, United States, gallons/minute
 hd = head of fluid, feet
 sg = specific gravity relative to water @ 60°F, 29.92 inches mercury
 n = overall pump/motor efficiency = 40%
 H = hours of operation

Agitator horsepower for the 20,000 gallon tank is assumed to be 3 horsepower. An electricity cost of \$0.05 per kilowatt-hour (Reference 9) was used in the power costs determination. Unit cost and quantity of nitrogen needed for blending the blend tank are used on data taken from Reference 6.

Steam is needed for the thermal pretreatment. An average cost of \$6.00/1,000 pounds of steam at 100 pounds/square inch was used. The indirect operating costs include the costs of overhead, taxes, insurance, administration expenses, and capital charges. Taxes, insurance, and administration can collectively be estimated at 4% of the capital costs, while overhead charges can be considered at 80% of the labor charges for both O&M. The annualized capital charges reflect the costs associated with capital recovery over the depreciable life of the system and can be determined as follows:

$$\text{Capital Recovery Cost} = (\text{capital costs}) \times \frac{i (1 + i)^n}{(1 + i)^n - 1}$$

where: i = annual interest rate
 n = capital recovery period

For these estimates, a useful life of 15 years and an average annual interest rate of 10% were assumed.

Blending O&M Costs

The estimated annual O&M costs for the previously described tank blending system is \$36,000. These costs are based on a five day per week blending operation using a 20,000 gallon blend tank equipped with agitation and nitrogen blanketing. Wastes are either transferred from other parts of the plant or unloaded from a tank truck or rail car into the tank where they are upgraded to the desired quality by adding other wastes or fuel oil. The waste mixture is continuously agitated to achieve uniform blending. After the desired quality is obtained, the blended waste is pumped to either another tank for long-term storage or directly to the burner. Costs are based on the assumption that it takes eight hours to prepare 20,000 gallons of blended waste which is also the maximum daily production rate of the system. One operator is required two hours per day to oversee the blending process. Utilities include electricity to drive the agitator and transfer pumps and nitrogen for blanketing the tank. The nitrogen can be either vented to the furnace or to the same system as used for the storage tanks.

The blending O&M costs will vary with the amount of fuel blended. Costs presented here are for one 20,000 gallon tank system which will provide 595 gallons per hour of blended waste. O&M costs for furnaces firing more than the capacity of the 595 gal/hr blending system can be obtained by multiplying

the costs presented here for this system by the number of multiple blending systems required. This can be determined by dividing the production capacity to be fueled with waste by 595 gal/hr and rounding to the nearest whole number. For example, a 3,000 tons/day blast furnace firing 7 gallons of waste fuel per ton pig iron produce would consume 875 gal/hr of the waste. Two 20,000 gallon blending systems (595 gal/hr) are required to produce the necessary waste fuel.

$$\frac{875 \text{ gal/hr}}{595 \text{ gal/hr}} = 1.47 \text{ (2 rounded to nearest whole number)}$$

Therefore, the O&M costs to blend the waste fuel for this furnace is two times the costs presented for the 20,000 gallon system, i.e., \$72,000.

Straining O&M Costs

The O&M costs associated with removing grit and debris with basket strainers are considered insignificant for purposes of the RIA.

Thermal Treatment O&M Costs

Figure 5.3 presents the annualized thermal treatment O&M costs as a function of waste feed rate for the two categories of waste for which capital costs were given earlier in this section:

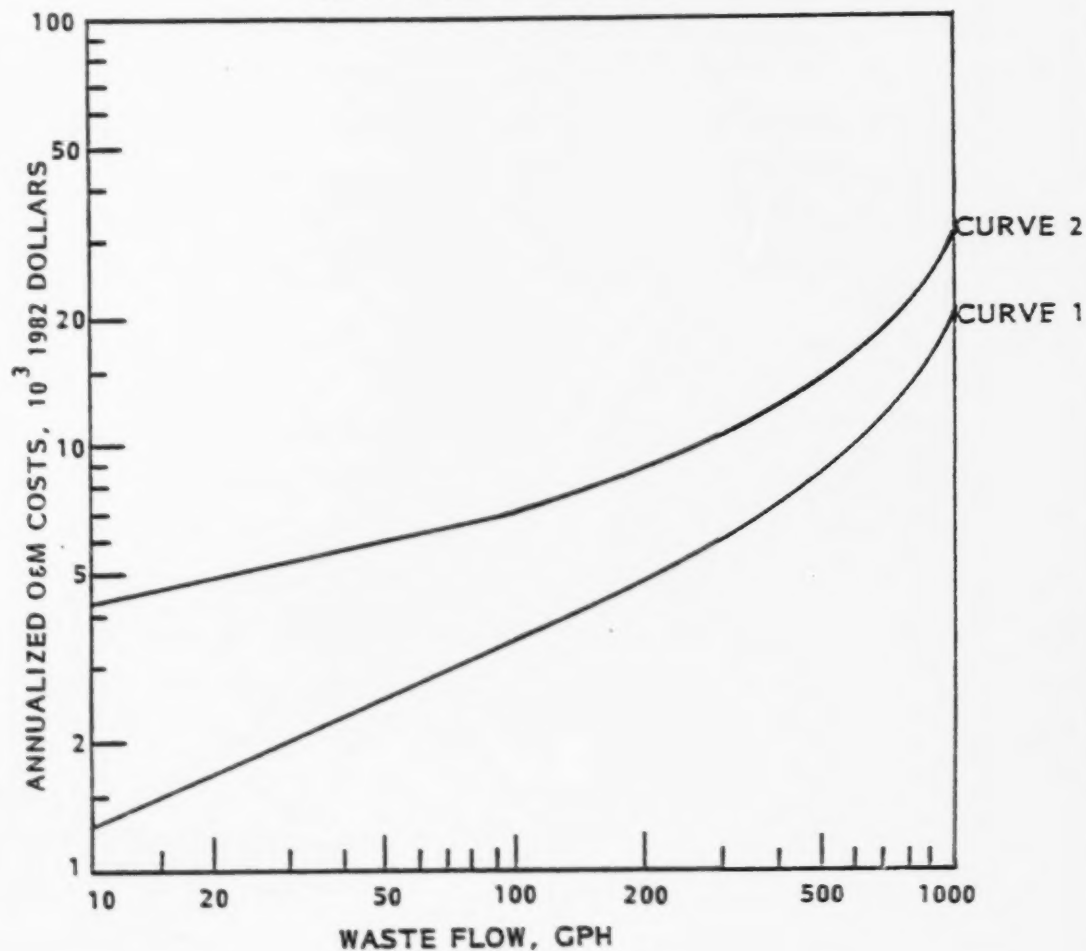
- o Curve 1 - Wastes with viscosities greater than 750 standard saybolt units but less than 10,000 standard saybolt units at ambient temperatures.
- o Curve 2 - Wastes with viscosities greater than 10,000 standard saybolt units at ambient temperatures.

Provisions for heating and insulating of the storage tanks are not included in the O&M costs presented in Figure 5.3. Wastes with viscosities greater than 10,000 standard saybolt units require tank heating and insulation while wastes with viscosities between 750 and 10,000 standard saybolt units at ambient temperatures do not. The following are O&M costs for storage tank heating and insulating:

O&M Costs - Thermal Treatment

<u>Tank Size (gallons)</u>	<u>O&M Costs (1982 dollars/year)</u>
500	470
1,000	890
2,000	1,710
5,000	2,190
10,000	3,900
20,000	5,380

FIGURE 5.3

ANNUALIZED O&M COSTS FOR THERMAL TREATMENT
OF HAZARDOUS WASTES

CURVE-1- PUMPABLE AT AMBIENT TEMPERATURES BUT MUST BE HEATED FOR GOOD ATOMIZATION.

CURVE-2- MUST BE HEATED TO KEEP PUMPABLE AND TO OBTAIN GOOD ATOMIZATION.

Furnace Modification O&M Costs

Burner Modification O&M Costs

The burner gun to fire waste fuel should require little operator attention beyond that required for the furnace firing conventional fuels. Maintenance cost will be about 5% of capital cost, capital recovery 13.2%, and taxes, insurance, and administration, 4%. Total annual O&M cost will be about 22.2% of the original burner gun assembly capital cost.

Oxygen and Carbon Monoxide Monitoring O&M Costs

Annual operating costs for several continuous monitoring systems have been estimated for evaluating the economic impact of NSPS for industrial boilers (Reference 7). These costs should also be applicable to "clean" exhaust gas applications. An annual O&M cost of \$18,500 in 1982 dollars was estimated for an oxygen monitoring system. Costs for a carbon monoxide system should be about the same. The major items included in this estimate are the maintenance and performance certification. One-half manhour per day was assumed to be required for the maintenance at a rate of \$35.81/manhour, including supervisor and overhead. One certification test per year, costing \$11,900 was assumed.

The operating and maintenance cost of a CEMS for dirty gas is expected to be higher than for a clean gas application. The extent to which these cost will be higher will depend on the type of conditioning system selected and to some degree will be site-specific. The labor will be approximately 50% higher for a system for a dirty gas application based on an assumed one-half manhour per day labor requirement. Thus, for a dirty gas system, three quarter-manhours per day would be required for O/M. At the assumed \$35.81/manhour labor rate (1982 dollars) used above the O/M cost would be \$21,000 (1982 dollars) which includes \$11,900 for one certification test.

Requiring O₂ and CO monitors on industrial furnaces burning hazardous waste could result in a fuel cost savings to the operators of these devices. This potential saving would result if the operators used the O₂ monitors to maintain low excess air (LEA) combustion of the fuels. With LEA combustion, less fuel is required because less heat is lost out the stack with the combustion gases.

The magnitude of the potential fuel savings that can be obtained by LEA combustion must be determined individually for each furnace because it depends on many factors. The major factors influencing the potential savings include:

- o Furnace type and condition
- o Burner type and condition
- o Combustion control type and condition
- o Operating load level

The furnace type and condition have a large impact on the amount of fuel saving that may be achieved through LEA combustion. Some types have design characteristics that limit the range of LEA operation. Also, the flue gas exit temperature for one type furnace can be significantly different from those of another type. Since the fuel savings for a given excess air reduction is temperature dependent furnaces with higher exit flue gas temperatures

should be capable of achieving a higher fuel savings per unit excess air reduction. The condition of the furnace also impacts the fuel savings potential of LEA. A furnace that has significant air in-leakage is more difficult to operate at low excess air levels because the air infiltration may distort the O₂ reading drastically.

The type and condition of the burner(s) installed in the furnace also greatly influence the fuel savings potential of LEA operation. A burner is designed to operate efficiently over a specific excess air range. If operated at an excess air range lower than the design level, proper mixing of the fuel and combustion air cannot be achieved. Poor air and fuel mixing would likely result in incomplete combustion of the fuel and higher fuel consumption. Gas burners generally can operate at lower excess air levels than oil burners which generally operate at lower levels than coal burners. There is also a wide variation in the excess air level operations capability of burners for a given fuel. For example, design excess air levels for oil burners used in asphalt dryers can vary from five to twenty percent. The condition of the burner also affects the potential fuel savings because the fuel flow through a dirty or damaged burner is difficult to control.

Another important factor determining the potential of LEA combustion and hence fuel savings is the type and condition of the combustion controls. Combustion controls vary widely in complexity from the simple single-point positioning units typically found on smaller units to the metering system of a complex, computerized process control system. Interfacing the O₂ monitor to these systems has limitations that are unique to each type of control system. The level of LEA achievable is limited to how well the O₂ monitoring is used by the control system. Also, the condition of control system mechanical components also impacts the fuel savings potential. Damper linkage may flex slightly, and bearings may wear over time. Even metering systems are susceptible to some shortcomings, since their flow transmitters are operated at temperatures and pressures that vary significantly from those at which the transmitters were initially calibrated.

More excess O₂ is needed at low loads because of poorer mixing of the fuel and air. Consequently, the operating load level also impacts the fuel savings of LEA combustion.

Because of the influence of the factors discussed above, a detailed breakdown of potential fuel savings by furnace type and fuel is deemed unjustified. For the purpose of determining the economic impact of requiring O₂/CO monitoring, two estimates of this potential fuel saving is considered adequate. A 2% savings is estimated as a typical average value for all the furnaces discussed in Section 3 except asphalt and lightweight aggregate kilns. A 4% savings is estimated for asphalt and lightweight aggregate furnaces because these units are typically operated at very high excess air levels. Both estimates are based on discussions with combustion control equipment suppliers, information found in the literature, and limited data on oxygen levels and furnace exit flue gas temperatures. These estimates are also based on the assumption that a 0.5% increase in combustion efficiency is achieved per 1% reduction in the O₂ level in the flue gas.

Not all furnace operators would receive the fuel saving estimated for LEA operation as a consequence of requiring O₂/CO monitoring. Many of the larger industrial boilers are already equipped with O₂-trim or O₂/CO-trim. In fact, those industrial furnaces equipped with CO-trim systems will actually be required to operate at a higher LEA level than they are currently operating at and thus would consume more fuel. Furnaces with CO-trim systems typically operated with a CO setpoint of from 200 to 400 ppm. A 100 ppm CO limit would require that they operate at a higher LEA level than their current setpoint level. Also, nearly all sulfur recovery furnaces are equipped with O₂-trim systems. Thus, estimating the fuel savings potential of requiring O₂/CO monitors on a particular population of furnaces requires a knowledge of how many units in the population already are operating at LEA levels (i.e., how many are using O₂ or CO monitors to achieve LEA combustion.) Once the fraction of furnaces currently employing LEA controls is determined, the potential fuel savings for the entire population may be estimated from the potential savings of a single unit.

The savings of a single furnace can be estimated by multiplying the total annual fuel cost by 0.02 or 0.04 depending on the type fuel burning device being considered. The annual fuel cost is estimated by multiplying the design heat impact by the unit fuel cost presented in the preceding paragraphs. For example, the maximum potential annual fuel savings of a 150 x 10⁶ Btu/hr, residual oil-fired furnace would be:

$$.02 \times 150 \times 10^6 \text{ Btu/hr} \times 8760 \text{ hrs/yr} \times \$4.62/10^6 \text{ Btu} = \$121,400$$

This assumes that the furnace operates 24 hours per day, 365 days a year. This savings can be adjusted to match different assumptions regarding load factor.

A comprehensive survey of furnaces was not conducted to determine the fraction of units already equipped with an LEA capacity. Conversations with industry representatives and control system vendors indicate the approximate breakdown listed below:

<u>Furnaces</u>	<u>% With Some Form of LEA Cap</u>
Lightweight Aggregate	30
Blast Furnaces	
Stoves	50
Boilers	50
Spent Acid Recovery	50
Asphalt	0
Reverberatory	50

Air Pollution Control Devices O&M Costs

The annualized O&M costs for the venturi/gas absorption systems presented in Figure 5.4 are based on 8,700 hours/year operating time and the cost factors presented in Table 5.7. The O&M costs include direct costs such as operating labor and materials, maintenance, replacement parts, utilities, and collected particulate disposal. Also included are indirect costs such as overhead, insurance, taxes, and capital recovery.

Estimated operating labor requirement for the air pollution control device system is four manhours per shift. The utilities include quench water, scrubber water, absorber water, caustic soda, and electricity.

FIGURE 5.4

ANNUALIZED OPERATION AND MAINTENANCE COSTS
FOR THE VENTURI/GAS ABSORPTION SYSTEMS

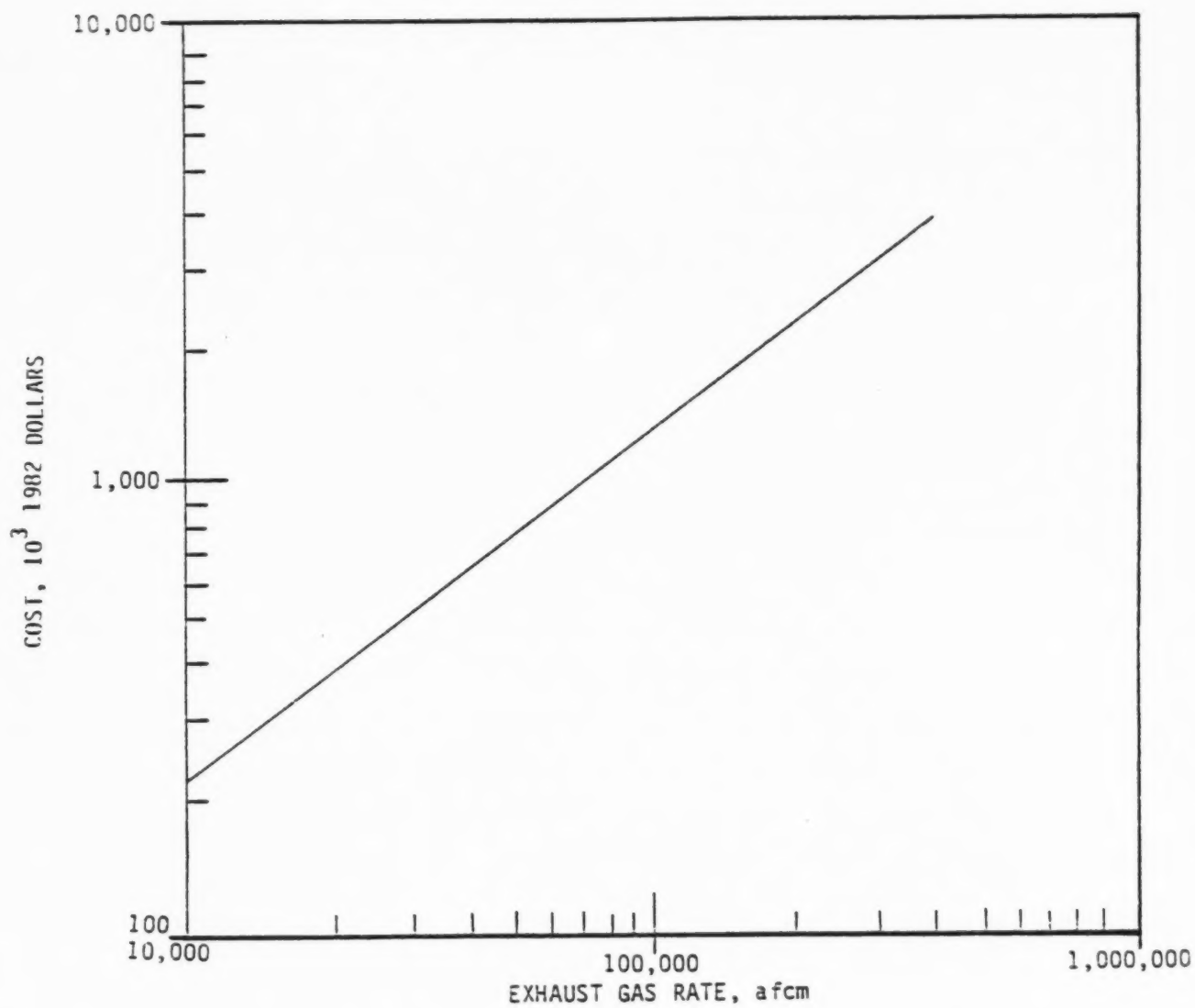


TABLE 5.7

COMPONENTS OF ANNUALIZED O&M COSTS FOR
AIR POLLUTION CONTROL DEVICES

Direct Operating Costs	Cost Factor ^a
Operating Labor	
- Operator	\$9.75/manhour
- Supervisor	15% of operator
Maintenance	5% of capital costs
Utilities	
- Electricity	\$0.5/kilowatt-hours
- Steam	\$6.00/1,000 pounds
Indirect Operating Costs	Cost Factor ^a
Overhead	80% of operating labor and main- tenance labor
Property Tax	1% of capital costs
Insurance	1% of capital costs
Administration	2% of capital costs
Capital Recovery Costs	0.132 (using $i =$ 10% and an equipment life of 15 years)

^a All costs are in 1982 dollars.

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